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PRESSURE FILTRATION OF WASTE WATER SLUDGE WITH ASH FILTER AID

James W. Gerlich, et al

Cedar Rapids Water Pollution Control Plant
Iowa

June 1973

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27 Abstract

Cedar Rapids, Iowa used pilot plant studies as an effective approach to an economic solution to dewatering secondary digested sludge. After piloting several dewatering processes, the pressure filter system was selected and a full scale plant was constructed. During the course of the dewatering studies it was observed that fly ash was an effective filter aid.

The full scale plant was designed to utilize sludge ash from incinerated filter cake, as well as power plant fly ash. The design capacity is 28 tons of dry sewage solids for 16 hour operation at 48 percent dry solids cake.

Performance data from the full scale plant ~~was~~ were evaluated over a period of approximately nine months. Both fly ash and sludge ash were evaluated as a filter aid, with and without chemicals. Economic evaluations were made of operation and equipment.

Some ~~of the~~ conclusions are: (1) Pressure filtration of waste water sludges is an effective and economical process. (2) Ash filter aid increases dewatering production and decreases chemical costs. Sewage sludge ash from an incinerator can be recycled to process; power plant fly ash is even more effective sludge conditioner. (3) A detailed pilot plant program is of great value in design of a full scale plant. (4) Some chemicals in combination with ash filter aid further improve dewatering efficiencies. (EPA)

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PRESSURE FILTRATION OF WASTE WATER SLUDGE WITH ASH FILTER AID

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ABSTRACT

Cedar Rapids, Iowa used pilot plant studies as an effective approach to an economic solution to dewatering secondary digested sludge. After piloting several dewatering processes, the pressure filter system was selected and a full scale plant was constructed as the first major installation in the United States. During the course of the dewatering studies it was observed that fly ash was an effective filter aid which cut chemical conditioning costs from about $20 per ton to $4 per ton dry solids.

The full scale plant was designed to utilize sludge ash from incinerated filter cake, as well as power plant fly ash. The design of this facility was based upon data collected in the operation of the pilot plant, which data at that time represented nearly all of the analytical data available on the operation of the pressure filter in the United States. The design capacity is 28 tons of dry sewage solids for 16 hour operation at 48 percent dry solids cake.

Performance data from the full scale plant was evaluated over a period of approximately nine months. This data indicates that the full scale plant is capable of operation at a greater capacity and efficiency than that projected from pilot plant data. Both fly ash and sludge ash were evaluated as a filter aid, with and without chemicals. Economic evaluations were made of operation and equipment.

This report was submitted in fulfillment of Project No. 11060 EZX under the partial sponsorship of the office of Research and Monitoring, United States Environmental Protection Agency.

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SECTION I

CONCLUSIONS

1. Pressure filtration of waste water sludges is an effective and economical process.

2. Ash filter aid increases dewatering production and decreases chemical costs.

3. Sludge ash available from incineration of sewage solids can be recycled to the process as an effective filter aid.

4. Fly ash from a coal burning power plant is a more effective ash admixture than is sludge ash, however, haul and ultimate disposal is an additional inconvenience.

5. A detailed pressure filter pilot plant program is of great value in preparing for the process design. Good pilot plant data is reproducible in a full scale plant which has been confirmed by this report. This data may also serve as an operating aid for trouble-shooting problems that may occur during operation of the full scale plant.

6. Some chemicals in combination with ash filter aid further improve dewatering efficiencies at Cedar Rapids. Chemicals alone without ash filter aid are relatively ineffective in the range of $20 per ton dry solids versus $4 to $6 per ton in combination with ash at design conditions.

7. The full scale plant performed at greater capacity and efficiency than was projected from the pilot plant studies.

8. The pressure filter is very effective in dewatering raw primary sludges. This program is based upon secondary digested sludges, however, considerable experimentation was done with raw primary sludge at the termination of the basic program. More experimentation should be done with raw primary, as well as raw secondary sludge dewatering, by an extension of this program.

9. Filtrate from the pressure filter is low in suspended solids and phosphates and can be recycled without creating a significant load to the treatment process.

SECTION II

RECOMMENDATIONS

1. This program covered dewatering digested solids. An extracurricular investigation of raw primary sludge was conducted which consisted of extensive bench work and 60 pressure filter runs. The results were extremely encouraging. A more detailed investigation of the performance of the pressure filter on raw sludge should be undertaken. This investigation should include raw primary, raw secondary and combined raw primary and secondary.

2. The physical and chemical mechanism of ash should be further studied. Some fly ash works much better than others. In some cases fly ash is a better conditioner than sludge ash, while in other studies, the exact opposite has been reported.

3. An optimization of sludge thickening as this relates to sludge dewatering in a pressure filter should be undertaken.

4. An expanded investigation of polymer sludge conditioning should be made to better define the limitations under pressure filtration. Considerable bench work was done at Cedar Rapids using many polymers. Some showed satisfactory sludge conditioning characteristics in preparation for filtration, but failed under actual pressure filtration. Favorable performance was experienced with polymer C-7 on the vacuum filter, as an example, but unfavorable experiences resulted when applied to the pressure filter. Apparently polymer experiences related to low differential pressure vacuum filters are not applicable to high differential pressure filters.

5. Some field work is necessary to measure the true demand of the pressure filter. Experimentation at Cedar Rapids would indicate that the true demand of the filter, particularly during early cake formation is not well understood. It has been reported that some European pressure filter processes slowly apply pressure and during initial cake formation limit the rate of conditioned sludge applied. Experience at Cedar Rapids indicated that better filter performance could be expected with extremely high initial rate of application.

6. A better definition of the value of precoat would be of considerable benefit as well as an evaluation of materials other than ash. The effect of quantity and quality of precoat material, along with the effect of pressure should be studied. The possibility of no precoat should also be explored in more detail.

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7. Deterioration of sludge conditioning in both the mix tank and the contact tank could have a direct relationship to filter performance. Some earlier bench work at Cedar Rapids indicated a sludge deterioration curve starting about 4 to 6 hours after conditioning, however, later bench work could not confirm this for the actual filter operation. It is believed this work could be directly related to improved design of the mix tank and contact tank sludge conditioning phase.

SECTION III

BACKGROUND AND INTRODUCTION

In 1966, the City of Cedar Rapids began testing ways of dewatering primary and secondary digested sludge from its Water Pollution Control Plant. At the outset of the dewatering investigations, laboratory work soon confirmed that the sludge from the Cedar Rapids plant would be difficult to dewater and that the filterability was unpredictable. Samples drawn from the digesters indicated that the sludge characteristics varied considerably. This was attributed to the complex and varied industrial wastes which entered the plant along with other wastes from the City. In terms of numbers, the BOD_5 population equivalent at the time of this program development was about 800,000 whereas the actual population was about 113,000.

To evaluate the sludge dewatering equipment available and to apply it to these complex wastes, numerous pilot plants were operated including two sludge centrifuges, a gravity (plug) filter, vacuum filter and a pressure filter. In addition, sludge samples were treated at a wet air oxidation process representative laboratory. Some of these programs were a near failure from the point of sludge conditioning and/or sludge dewatering. The centrifuge showed some promise in the percent of total solids discharged, however, the percent solids captured and the percent solids in the concentrate was very poor. Following the centrifuge pilot programs was a belt-type vacuum filter.

Preceding the vacuum filter pilot study, detailed laboratory work on digested sludges was undertaken using the Buchner funnel tests to obtain optimum chemical dosages for sludge conditioning. This work was followed by leaf tests to determine the best yield and most suitable filter media. These laboratory investigations indicated that the Cedar Rapids sludge would not dewater at any reasonable cost. Chemical costs alone were in the range of $20 per ton dry weight. Chemicals used were combinations of ferric chloride, lime and every polyelectrolyte available. Combinations of digested and raw sludge were also investigated and this did not affect performance.

In January 1967 experimentation with some conventional and unconventional filter aids was begun in an effort to reduce the chemical costs. Some of the admixtures investigated are:

1. Volcanic clay
2. Lime waste slurry from municipal water softening plant
3. Clay fines from gravel plant spoil pile
4. Waste activated carbon from a local industrial process
5. Pulverized recycled dried grit cake from waste treatment plant
6. Recycled grit and fibrous material from plant grit chambers
7. Fibrous wastes from packing industry
8. Commercial diatomaceous earth
9. Heated sludge to 140°F.
10. Fly ash from two power generation plants

Some of these materials were a complete failure. Others showed some promise, but only fly ash performed satisfactorily. The fly ash used was from a fossil fuel fired steam power generation plant where coal is fired as a pulverized fuel. Fly ash is in reality the total ash product of suspended pulverized fuel (rather than only the precipitated ash product of the stack as from many furnaces utilizing a traveling grate for the combustion area). Fly ash as a filter aid reduced the chemical costs to approximately $4 per ton dry weight (compared to $20 per ton) and delivered 28 percent cake solids from a vacuum filter. Details of the vacuum filter study with fly ash sludge conditioning appears later in this report.

In October 1967, a pressure filter was installed and operated for a five month period using fly ash as a filter aid. No chemical conditioning was used. The pressure filter may be generally described as a plate type consisting of 6 plates, 24 inches in diameter having approximately 36 square feet of filtration area, operated at a shut-off pressure of about 260 pounds per square inch (psi). The pressure filter has been in service in the United States for fifty plus years in industrial processes, but it has only been in recent years that its application to dewatering sewage sludge has been seriously considered. At this time the apparent attraction to the pressure filter was its success in Europe and England operating without the use of chemicals in numerous installations and its ability to produce a relatively dense sludge cake with extremely clear filtrate. Analytical data on the performance of the pressure filter was nearly non-existent at that time in this country and therefore the Cedar Rapids pilot program was undertaken with very limited guidelines for evaluating the process. After several early filter cloth failures, a cloth with suitable characteristics was obtained and considerable operational data was accumulated. It was recognized early during the pilot study that the pressure filter offered numerous advantages which qualified this process for consideration in the design of the proposed Water Pollution Control Plant improvements for Cedar Rapids. Details of the pressure filter pilot study and evaluation of a full scale process plant are included in this report.

SECTION IV

VACUUM FILTRATION PILOT PLANT STUDIES

Bench Scale - Vacuum Filter Leaf Test

The Cedar Rapids Water Pollution Control Plant is a two-stage high rate trickling filter plant with separate sludge digestion. Incoming wastes are high-strength from numerous wet industries such as meat packing, metals, wet milling, and grain processing. The treatment plant covers about 30 acres with all of the facilities under roof using thin-shell concrete domes. Flow averages about 22 million gallons per day with maximum treatment at 64 million gallons per day. The population equivalent of organic loadings is approximately 800,000 persons. The plant has been operating at a design overloaded condition for organic and solids quantities for several years.

The feasibility of using fly ash as a filter aid in dewatering secondary digested sludge at Cedar Rapids was confirmed through a sequence of laboratory bench work and pilot plant studies. The pilot studies using fly ash filter aid began with vacuum filtration and finished with pressure filtration. A full scale sludge dewatering plant, using pressure filtration was constructed as the result of these pilot studies.

The purpose of the first test was to study the feasibility of vacuum filtration of digested sludge from the second stage digester. A bench scale leaf test was made on a mixture of secondary digested sludge mixed with chemicals and fly ash. The bench scale test equipment consisted of a circular "leaf" with exactly 0.1 square foot area, upon which was supported a piece of appropriate filter cloth. The test leaf was connected through a filtrate receiver to a source of vacuum. During the test, the leaf was immersed in the sludge mixture and filtration carried out for a timed period. The leaf was then withdrawn from the sludge sample and permitted to drain under vacuum for another timed interval. The cake formed on the leaf was then taken off, and weighed both wet and after drying. By making individual runs at a series of filtration times and drying times, the relationship between cake formation time and yield was obtained. From the same set of data, the effect of drying times on the final moisture content was also established.

The sludge mixture consisted of 4000 ml of 5 percent by weight secondary digested sludge, mixed with 500 grams of fly ash, 100 ml of 10 percent by weight ferric chloride and 20 ml of 2 percent by weight C-7 polymer.

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Temperature was ambient, fly ash used had approximately 18 percent moisture content. Mild agitation was performed to maintain a uniformly suspended slurry. Results of this typical run No.3 are shown in Table I.

Filtration Results

The sludge mixture filtered fairly well, the cake discharged from the media cleanly. The filtrate contained suspended solids that rapidly settled (5 minutes) in a graduated cylinder. Figure No. 1 shows a correlation of the Form Filtration Rate vs. Cake Formation Time. This curve is from data obtained on test No. 3, Table I. Tests No. 1 and 2 were not plotted. On test No. 2 there was blinding and on test No. 1, the medium was too open. Ideally, the slope of the curve shown on Fig. No.1 should be -0.5. The plot shows -1.0. It is believed that the openings in the cake, after a short interval, were sealed off by the fines in the sludge mixture. Essentially, the weight of the cake was the same for all form times, because of this sealing off process. This doesn't indicate that the mixture is not filterable but that faster cycle time should be used, as the cake formation occurs only for a short period of time for this sludge mixture.

TABLE I

Test Run No. 3 - Digester No. 7
Vacuum Filter 0.1 sq.ft., leaf type
Cloth - Nytex 413
Sludge Solids - 5%
Temperature - ambient

4000 ml. sludge
500 grams wet fly ash
410 grams dry fly ash
100 ml. $FeCl_3$ at 10%
20 ml. polymer C-7 at 2%

Test No.	% Solids Feed	Filtering Vac.In.H_g Form	Dry	Filtering Time Min. Form	Dry	Filtrate Ml.	Gal/hr./sq.ft.	Cake Thickness	Cake Wt. Grams Wet	Dry	Percent Moisture	Dry Sludge Solids Lbs/Hr/Sq.Ft.
1	10.8	25.5	24	.5	3	161	51.3	1/4	64.2	24.2	62.3	20.5
2	10.6	26	25	1	2	177	28.0	1/4	68.4	25.9	62.1	11.1
3	11.5	25	24	2	1	174	13.8	5/16	91.8	30.5	66.8	6.1
4	11.1	24.5	24	1	1	143	22.6	3/16	65.0	23.2	64.3	9.8
5	11.0	24	24	3	.5	168	8.8	3/8	98.6	29.4	70.2	4.2
6						Void Run						
7	10.8	24.5	24	1.5	1.5	139	1.7	3/16	58.4	21.2	63.8	6.0

Figure No. 2 is a plot of Form Filtrate Rate vs. Form Time and the curve was obtained in a manner similar to the plot for Figure No. 1. The slope of this line is the same as the slope of the Form Filtration curve.

Figure No. 3 is a plot of the Full Scale Filtration Rate vs. Cycle Time. This curve was obtained from Figure No. 1. Using a 0.8 scale-up factor and air effective submergence of the filter drum of 25 percent, a point on the full scale filtration curve was obtained as follows:

Form time = one (1) minute
Yield at one (1) minute form time = 38#/hr/sq.ft.

At 25 percent effective submergence, 25 percent of the filter area is available to form a filter cake, therefore, the cycle time is:

$$\frac{1 \text{ minute}}{0.25} = 4 \text{ minutes per revolution (MPR)}$$

Full scale yield = 38 x 0.25 x 0.8 = 7.6#/hr/sq.ft.

The slope of the full scale filtration curve is the same as the slope on the form filtration rate curve Figure No. 1.

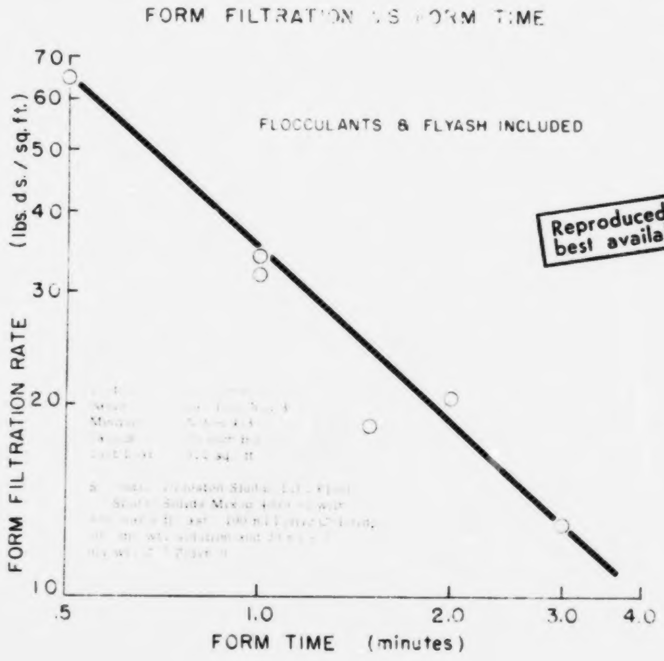

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Figure 1

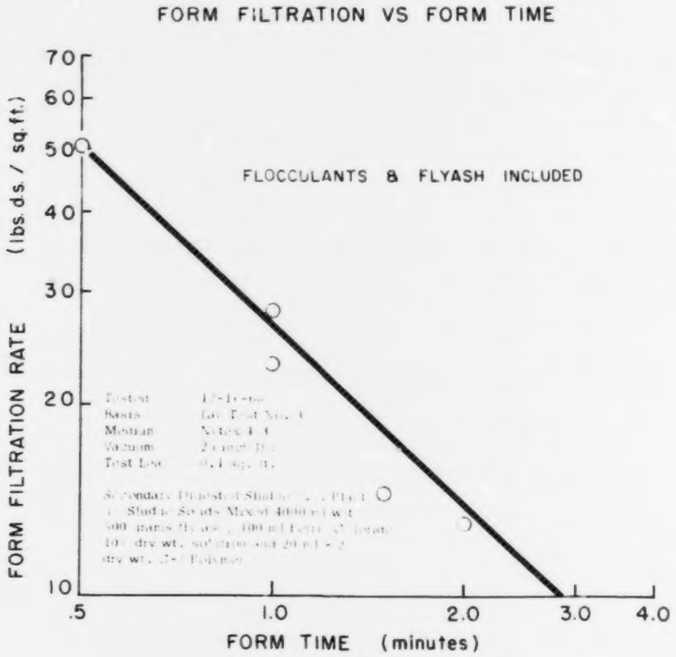

Figure 2

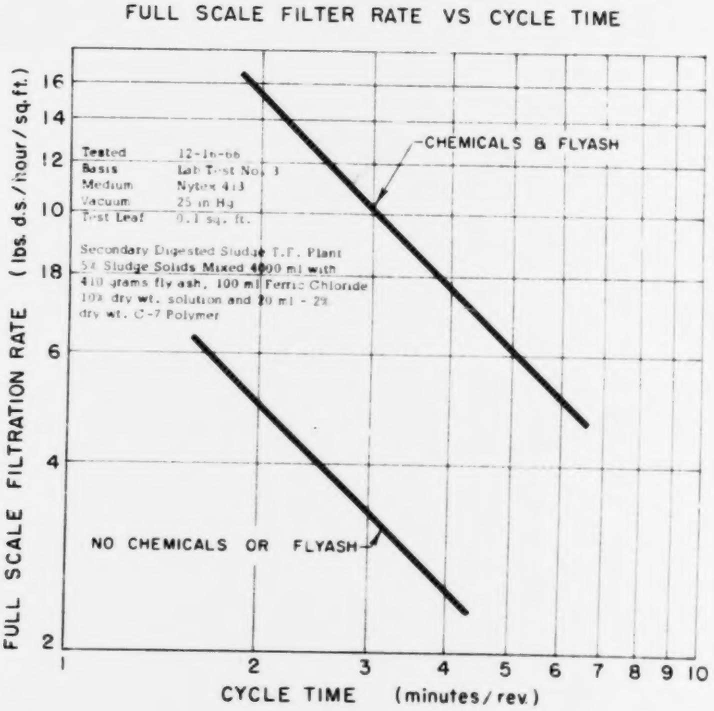

Figure 3

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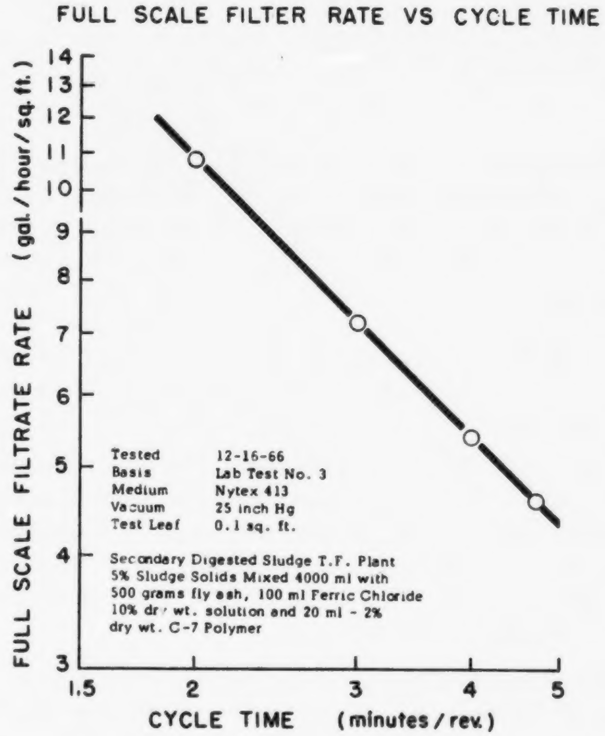
In order to size the filter station, it is necessary to know the full scale filtration rate of the secondary digested sludge. This was obtained by calculating the proportion by weight of the digested sludge in the mixture and assuming the filter cake followed the same weight proportion. This was calculated as follows:

4000 ml 5% by weight sludge	200 grams
410 grams fly ash	410 grams
100 ml $FeCl_3$ 10% by weight	10 grams
	620 grams

$$\% \text{ digested sludge} = \frac{200}{620} = 32\%$$

The secondary digested sludge without additives was then plotted as 32 percent of the secondary digested sludge with additives.

The Full Scale Filtrate Rate Curve (Figure No. 4) was calculated and plotted in a manner similar to Figure No. 3. The purpose of this curve is to estimate the filtrate flow.



Tested 12-16-66
Basis Lab Test No. 3
Medium Nytex 413
Vacuum 25 inch Hg
Test Leaf 0.1 sq. ft.

Secondary Digested Sludge T.F. Plant
5% Sludge Solids Mixed 4000 ml with
500 grams fly ash, 100 ml Ferric Chloride
10% dry wt. solution and 20 ml - 2%
dry wt. C-7 Polymer

Figure 4

Figure No. 5, Cake Moisture vs. Correlating Factor is a plot of percent cake moisture versus drying time divided by cake weight per unit area. A good correlation was obtained. The purpose of this curve is to determine optimum moisture content relative to operating conditions.

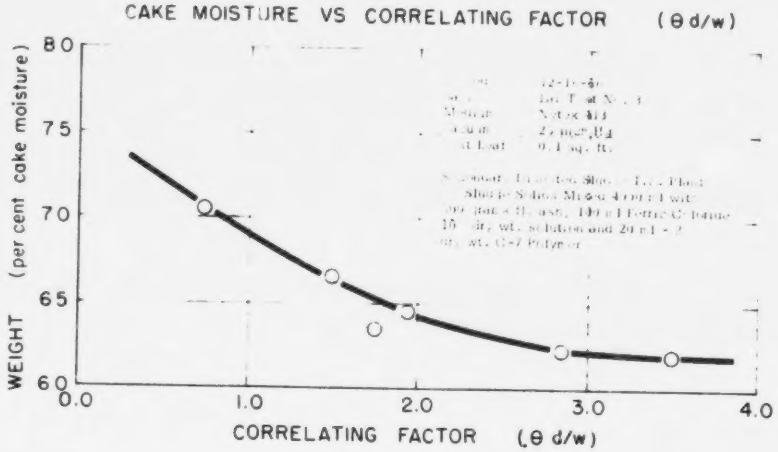


Figure 5

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Summary

1. The mixture of digested sludge, fly ash and flocculants filtered very well, giving a good discharge, with filter yields ranging to 20 lbs. per hour per square foot for short form times. The cake had a comparatively low moisture content ranging from 62 to 70 percent. The filtrate appeared to be fairly high in suspended solids by observation. It was suggested that these fines be separated from the filtrate before the filtrate is returned to the treatment system. The purpose of this is to prevent any possible build up of non-separable substance in the treatment system. The material in the filtrate settled out very rapidly.

2. Full scale filtration of the sludge mixture tested is feasible. It is our opinion that further leaf tests be run before a pilot plant study is made. When optimum leaf test filtration results are obtained, based upon an economic study of the various factors, then a short term pilot plant study would be feasible. Further leaf test studies of the fines in the sludge mixture should be made.

3. The fines in the sludge mixture caused suspended solids in the filtrate and while not detrimental to filtration, limited the cake formation by sealing off the openings in the cake. Further leaf tests on varying dosages of fly ash and flocculants might indicate ways to lessen the effect of the fines.

4. It was noted that on Test 1 data sheet, using Nytex 415, the back calculated feed solids are about 8 percent, Test 3 data sheet, Table I, using Nytex 413, the back calculated feed solids are about 11 percent. The Nytex 413 has a slightly closer weave than the Nytex 415. This would indicate the effect of fines or the size distribution of the fines.

Percent solids in the sludge admixture was probably 15 or 16 percent. The Nytex 413 captured more of the fines than the Nytex 415 as indicated by the back solids feed calculation.

5. It was not determined what proportion of fines in the filtrate originated from the fly ash conditioner or the sewage sludge.

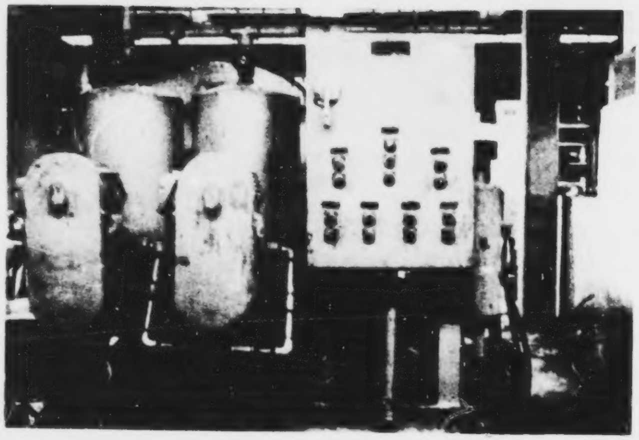
6. We believe that the filtration tests, as reported herein, were well worth running. The material is filterable but further bench scale tests are in order to evaluate certain points. This work was done in conjunction with the pilot plant study.

Pilot Scale Vacuum Filter

Following the bench scale vacuum filtration leaf testing program reported in the previous section of this report, a pilot scale vacuum filter was obtained and operated for a period of 10 weeks. The basic objectives of the pilot scale vacuum filter program were:

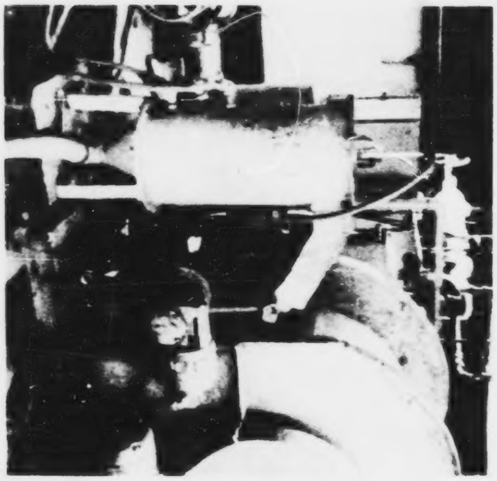
1. To determine the filterability of digested sludge at Cedar Rapids.
2. To determine the cost of chemicals and additives to filter digested sludge.
3. To establish the fly ash to sludge ratio for optimum filtration.

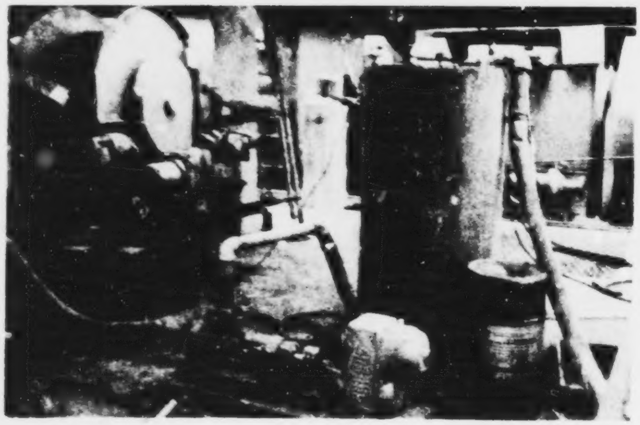
The equipment used is shown in Figure 6, and the flow schematic in Figure 7. The vacuum filter was a continuous belt filter, 3 ft. diameter by 1 ft. face. Two nylon belts were selected from previous laboratory evaluations. Both were monofilament nylon. One was a fine weave designated by the manufacturer as Nytex 413 . The other a coarser weave, designated as Nytex 415. The Nytex 413 was used exclusively during the test run as a convenience to data evaluation. Other filter media may warrant consideration.



Vacuum and Filtrate Pumps and Receivers Operating Controls

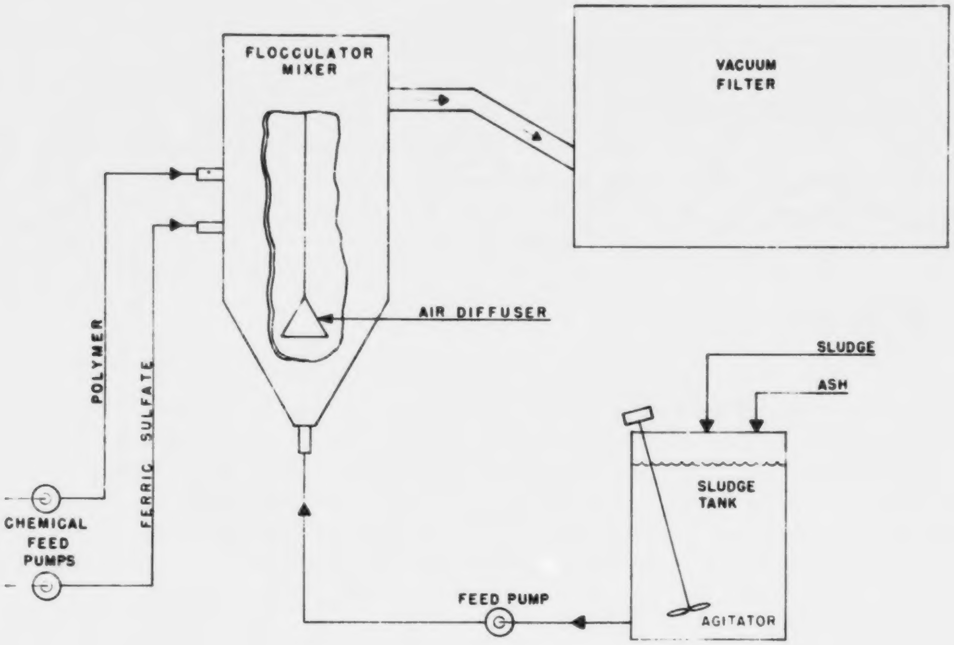
Flocculator Mixer





Sludge Mixing Tank and Variable Speed Feed Pump

Figure 6


FLOCCULATOR MIXER
VACUUM FILTER
AIR DIFFUSER
SLUDGE
ASH
POLYMER
FERRIC SULFATE
SLUDGE TANK
CHEMICAL FEED PUMPS
FEED PUMP
AGITATOR

SCHEMATIC DIAGRAM OF PILOT PLANT

Figure 7

The sludge pump was a rotating cavity positive displacement pump, with a variable speed motor drive. The pump was calibrated both by measuring the displacement at the discharge side and the drawdown in the 225 gallon sludge tank. The displacement of the pump at the setting used (No. 2) was 4.00 ± 0.05 gallons per minute.

Chemicals were added by positive displacement piston type pumps with variable stroke.

Chemicals and sludge were mixed in a flocculating tank by means of an air diffuser. This method proved very satisfactory and causes the least amount of shearing of floc while still maintaining adequate mixing.

Filtration Aids and Chemical Conditions

Fly Ash

The addition of fly ash to the sludge presented a problem and various methods were used in an attempt to prepare a homogeneous mixture of digested sludge and fly ash. The last method used, proved to be the most successful. This method involved screening the fly ash through a coarse mesh window screen over the vortex described by the mixer in the 250 gallon mixing tank. Although this method has been most successful, a combination of this coarser mesh screen and a finer mesh probably would have yielded better results. The fly ash used consisted of many basic oxides which upon addition to the digested sludge raised the pH to approximately 8.4. Since the fly ash was stored out-of-doors, the moisture content varied from 15 to 35 percent.

Polymeric Flocculants

The flocculant used in this test was a cationic polymer marketed by Rohm and Haas as "Primafloc C-7". This polymer had demonstrated superior effectiveness as a flocculating agent in the laboratory over other polymers used. Other polymers are now available which may be more effective or as effective as the one used, but considering the number of parameters which needed to be evaluated, it was felt that the introduction of another parameter would not be justifiable considering the length of time the pilot plant would be available. The polymer, a dry powder, was dissolved in water before use. Stock solutions of 2 percent by weight, were prepared. Dilutions of 0.6, 0.4, 0.3, 0.2 and 0.1 percent were prepared from this solution for introduction into the sludge. It was noted that the dilutions of polymer in the range of 0.1 to 0.01 percent are more desirable and full scale facilities should include chemical pumps capable of delivering the needed quantities of polymer at these lower dilutions.

Other Chemical Additives

Ferric sulfate was used to a lesser extent in this study than the polymer and was always used in conjunction with the polymeric flocculant. The apparent reason for using ferric sulfate was to lower the dosages of polymer needed to form an effective floc and not to replace it. Ferric sulfate is not usually considered to be as effective as ferric chloride, however, at increased dosages, it is still more economical considering the initial cost of 3 cents per pound versus approximately 10 cents per pound for ferric chloride.

Ferric sulfate was purchased as "Ferrifloc", a dry powder, and dissolved in water to form a 5 percent by weight solution, which was also the concentration used as the feed solution. Both ferric sulfate and the cationic polymer form acidic solutions and all tanks and feed lines, and fittings, should be coated to protect the metal. Sodium hydroxide was added to the sludge on one occasion, to determine the dependence of sludge filtration on pH. The pH which was 8.4 was raised to 9.4 and filtration immediately ceased, that is, no dischargeable cake was obtained.

Test Procedure

The vacuum filtration was performed on a batch basis. The holding tank held approximately 225 gallons of sludge, which was enough for a 56 min. run at a constant four gallons per minute feed rate. The fly ash was added immediately and the run was started. The time between the drawing of the sludge to the beginning of the run was never more than one hour. During the run, the cycle time was varied from approximately one minute to four minutes depending upon cake dischargeability. Each cycle time was timed by a stop watch. Samples of the filter cake were taken using a ring exactly 0.1 square foot of area. The samples were then dried in the laboratory at 102° C. to determine dry weight. The dry weight in grams was then converted to pounds per square foot per hour by the following formula:

$$\frac{\text{Cake Wt. in Grams}}{\text{453.6 Grams/Lb.}} \times 10 \text{ sq.ft.} \times \frac{\text{60 Min/Hr.}}{\text{Min./Cycle}} = \text{lbs./sq.ft./hr.}$$

Many duplicate samples were taken at the same cycle time. This was done because at times, the filter cake was not uniform across the filter media. This condition is aggravated **by** the fact the filter face is only one foot long, secondly, the digested sludge contained bits of matter not representative of the whole which could give erroneous weights for a particular cycle time. Finally this method gives the operator a good estimate of the consistency of operation and the precision of his technique.

Summary - Operating Procedure
General

Cloth	Nytex 413
Length of Run	56 minutes at 4 gallons per minute, sludge wasted over overflow weir on filter.
Condition of Cloth	Good - three small holes developed during last week of run.
Sludge	4 to 7% solids, 40-50% volatility pH 7.0 ± 0.2 Rate of filter feed - 4 gallons per min.
Filtering Minutes Per Revolution	0.73-6.00 depending upon cake quality
Vacuum	22-25 inches Hg

Percent Submergence	25% constant
Number of Wash Headers Used	5
Chemicals	Polymer C-7 only, dilution 0.1 to 0.6% Ferric Sulfate 5% feed (when used) Fly Ash Ratio 0.25 to 1.90 sludge Fly Ash 15-35% moisture
Filtrate Clarity	The suspended solids in the filtrate ranged from 0.2% solids to 0.64% with the average very nearly 0.5% solids (5000 mg/l).

Cake thicknesses were measured by the metric system and the English system. The thickness of the cake is not of great importance in itself, because of its dependence upon moisture content and fly ash concentration.

Explanation of Data Sheet

Date	Date of filter run
Percent Solids	Dry solids (at 102°C.) content of digested sludge
Fly Ash to Sludge Ratio	Ratios of weight of dry solids of fly ash to weight of dry solids of digested sludge.
Lbs./Ton Polymer	Pounds of polymeric flocculant per ton dry weight digested sludge solids only.
Lbs./Ton Ferric Sulfate	Pounds of ferric sulfate per ton dry weight digested sludge solids only.
Percent Moisture	Percent moisture of filter cake.
Lbs/Hr/Ft.^{2}/Total	Filter yield of all solids including fly ash
Lbs/Hr/Ft.^{2}/Sludge	Filter yield in terms of sludge only (fly ash not included)

Date 1967	Sludge % Solids	Fly Ash to Sludge Ratio	Pounds/Ton Polymer	Pounds/Ton $Fe_2(SO_4)_3$	Cycle Time	% Moisture	Lbs/hr/ft^2 Total	Lbs/hr/ft^2 Sludge	Discharge	Thickness Inches	Dollar Cost
4-12	7.71	0.706	6.90	-	4.00	65	4.5	2.63	C		6.90
					2.62	67	6.52	3.82	C		
					1.58	68	10.96	6.42	C		
4-18	7.40	0.253	11.00	-	6.40	77	7.12	5.69	B	1/2	11.00
					3.75	76	8.19	6.55	B	3/8	
					2.75	72	9.05	7.24	C	3/[illegible]	
					1.66	76	13.21	10.57	C	-	
					1.16	76	11.44	9.15	C	-	
4-18	7.74	0.359	9.33	-	6.50	75	4.46	3.27	B	1/4	9.33
					4.92	75	4.69	3.45	B	1/8	
					2.67	73	4.84	3.55	C	3/16	
					1.42	73	10.10	7.42	C	1/8	
5-17	6.40	0.531	8.64	44	7.00	72	6.64	4.33	A	1/2	9.96
					1.71	72	9.36	6.11	A	1/4	
5-25	6.30	0.974	4.34	44	6.50	64	4.02	2.03	A	13/64	5.66
					3.93	64	5.04	2.55	B	5/32	
					2.62	62	5.47	2.77	B	1/8	
					2.06	63	7.77	3.93	B	1/8	
					1.66	61	7.92	4.01	C	7/64	
5-31	6.90	0.760	2.16	54	3.92	66	2.17	1.23	B	9/64	2.78
					2.75	57	2.61	1.48	B	7/64	
					2.61	59	2.97	1.68	B	7/64	
					2.16	58	3.18	1.80	C	7/64	
					1.66	60	4.21	2.39	C	5/64	
6-13	5.80	1.00	4.76	80	3.92	67	6.13	3.06	B	7/32	7.16
					2.50	68	7.97	3.98	B	13/64	
					1.83	66	9.97	4.98	B	5/32	
					1.50	66	10.80	5.40	B	1/8	
					1.50	66	10.80	5.40	B	1/8	
6-15	5.75	1.412	3.12	64.3	3.70	-	7.69	3.18	B	11/64	5.04
					2.50	-	8.52	3.52	B	1/8	
					2.50	-	9.52	3.94	B	1/8	
					1.61	-	11.80	4.88	B	7/64	
					1.61	-	13.76	5.69	B	7/64	
					1.40	-	15.40	6.37	B	5/64	
					1.10	-	15.71	6.50	B	5/64	
6-15	5.75	1.42	3.93	64.3	3.58	-	8.78	3.63	B	5/32	5.85
					2.25	-	10.46	4.33	B	1/8	
					1.93	-	12.10	5.01	B	1/8	
					1.53	-	11.76	4.86	B	7/64	
					1.25	-	16.00	6.62	B	7/64	
6-15	5.70	0.96	5.88	48.3	4.08	70	6.11	3.12	B	7/32	7.33
					4.08	71	6.30	3.22	B	7/32	
					3.76	71	5.95	3.04	C		
					3.03	68	5.74	2.93	C	5/32	
					2.75	72	7.69	3.93	C	11/64	
					2.75	72	7.80	3.98	C	11/64	
					1.92	67	8.08	4.13	C	9/64	
					1.92	68	9.07	4.63	C	9/64	
					1.63	67	11.15	5.87	D	1/8	
					1.32	70	10.46	5.35	D	1/8	
6-21	5.0	1.55	5.51	-	3.05	57	6.27	2.45	B		5.51
					3.05	57	6.02	2.35	B		
					2.13	55	6.80	2.66	B		
					2.13	62	6.62	2.59	B		
					1.72	59	10.50	4.12	C		
					1.46	57	12.80	5.01	C		
					1.46	57	12.47	4.88	C		
6-21	5.0	1.55	3.67		3.91	55	3.82	1.49	B		3.67
					3.91	55	3.63	1.42	B		
					3.18	54	4.81	1.88	C		
					3.18	52	4.73	1.85	C		
					2.15	62	4.76	1.86	C		
					2.15	53	6.67	2.61	C		
					1.43	54	8.91	3.49	C		
					1.18	41	9.51	3.72	C		

Date 1967	Sludge % Solids	Fly Ash to Sludge Ratio	Pounds/Ton Polymer	Pounds/Ton $Fe_2(SO_4)_3$	Cycle Time	% Moisture	Lbs/hr/ft^2 Total	Lbs/hr/ft^2 Sludge	Discharge	Thickness Inches	Dollar Cost
6-22	6.6	0.951	4.17		3.05	58	4.72	2.42	B	1/8	4.17
					3.05	60	4.44	2.27	B	1/8	
					1.43	59	10.31	5.28	B	7/64	
					1.43	64	9.26	4.75	B	7/64	
					1.26	60	12.9	6.62	B	5/64	
					1.03	58	14.3	7.33	B	5/64	
					1.03	58	14.9	7.64	B	5/64	
					0.83	58	14.86	7.60	B	1/16	
					0.83	58	14.62	7.49	B	1/16	
					0.73	59	15.22	7.79	B	3/64	
					0.73	59	15.25	7.82	B	3/64	
6-28	5.20	1.02	6.93	-	3.20	66	8.46	4.23	B	5/32	6.93
					3.20	67	8.27	4.13	B	5/32	
					2.45	65	9.76	4.88	B	5/32	
					2.20	56	10.42	5.12	B	5/32	
					2.20	62	9.37	4.63	B	5/32	
					2.13	63	10.31	5.15	B	1/8	
					1.50	62	13.[illegible]	6.84	C	3/32	
					1.45	62	11.19	5.59	C	3/32	
6-29	6.6	1.07	4.09		3.07	59	6.79	3.27	B	5/32	4.09
					3.07	59	6.97	3.36	B	5/32	
					2.17	57	8.72	4.21	C	3/32	
					2.17	57	9.13	4.40	C	3/32	
					1.47	60	10.15	4.90	C	1/16	
					1.47	59	10.64	5.13	C	1/16	
					1.47	60	11.70	5.65	C	1/16	
					1.27	59	11.89	5.74	C	1/16	
6-29	6.6	1.07	2.73		3.08	55	3.95	1.90	C	1/16	2.73
					3.08	52	3.95	1.90	C	1/16	
					2.13	54	[illegible].21	2.51	C		
					1.45	-	No Discharge				
6-30	4.0	1.89	8.03		3.03	58	6.36	2.20	B	1/8	8.03
					3.05	58	6.49	2.24	B	1/8	
					2.10	57	8.08	2.79	B	3/16	
					2.10	59	7.84	2.71	B	3/16	
					1.43	54	12.80	4.42	C	3/16	
					1.43	58	11.30	3.91	C	3/16	
					1.03	59	13.28	4.59	C	1/16	
					1.03	56	13.53	4.68	C	1/16	
6-30	4.0	1.89	5.35		3.05	56	4.34	1.50	B	1/16	5.35
					3.05	56	4.68	1.62	B	1/16	
					2.10	46	5.74	1.98	C	1/16	
					1.46	56	5.91	2.04	D		
					1.46	-	5.43	1.88	D		
7-3	5.0	1.20	6.34	-	4.43	60	4.31	1.96	B	1/8	6.34
					2.52	61	6.83	3.10	B	1/8	
					2.16	63	7.76	3.53	B	3/32	
					2.16	61	7.53	3.42	B	3/32	
					1.50	61	9.79	4.45	B	3/32	
					1.50	57	10.99	5.00	B	3/32	
					1.07	62	10.88	4.95	C	3/64	
					1.07	63	11.93	5.42	C	3/64	
7-5	5.4	1.36	5.00	-	3.00	55	6.75	2.86	B	None	5.00
					3.00	51	6.67	2.82	B	None	
					2.17	53	8.21	3.48	B	None	
					2.17	52	8.27	3.50	B	None	
					1.50	53	11.27	4.77	B	None	
					1.50	53	11.20	4.74	B	None	
					1.03	51	13.4	5.68	B	None	
					1.03	51	13.7	5.80	B	None	

Discharge	Since cake dischargeability is a critical factor in filtration studies, cake discharge was given a rating from A to F and entered on the data sheets of the valid filter runs.
A	Very good discharge - thick cake
B	Very good discharge
C	Satisfactory discharge, clean cloth
D	Incomplete discharge
F	No discharge
Thickness	Thickness of filter cake in millimeters or inches , 25.4 millimeters = 1 inch
Chemical Costs	Costs were evaluated on the following: $1.00 per pound - Polymer $0.03 per pound - Ferric Sulfate

Test Results

Fly Ash to Sludge Ratio 1/1

Figure 8 illustrates the results of four runs with an approximate ratio of 1/1 fly ash to sludge. The actual conditions are:

Line	Date	Fly Ash Sludge Ratio	Ferric Sulfate Lbs /Ton	Polymer Lbs/ Ton	Chemical Cost/Ton
1	6-29	1.02	-	4.09	$4.09
2	6-28	1.02	-	6.93	6.93
3	6-13	1.00	80	4.76	7.16
4	5-25	0.974	44	4.34	5.66

The effectiveness of ferric sulfate is not apparent. These plots show the range of yields and costs which may be found over a period of time.

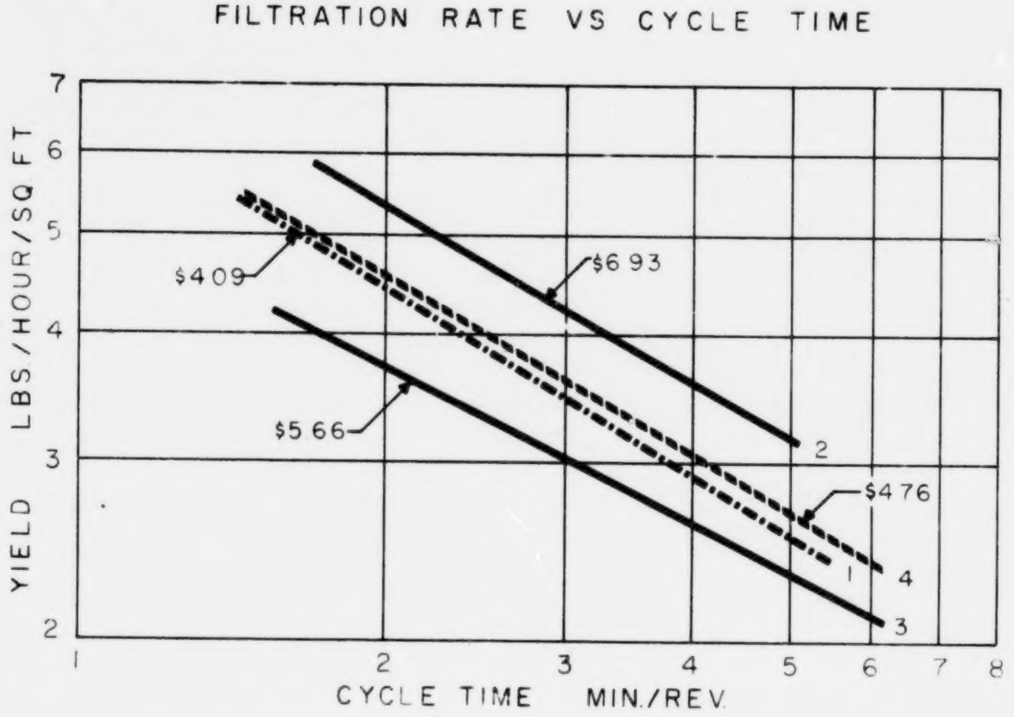


Figure 8

Fly Ash to Sludge Ratio 1.4/1

Figure 9 shows the result of four runs where the fly ash to sludge ratio was approximately 1.4 to 1. The actual operating dosages are given in the following table.

Line	Date	Fly Ash/Sludge Ratio	Ferric Sulfate Lbs/Ton	Polymer Lbs/Ton	Cost
1	7-5	1.36	-	5.00	$5.00
2	6-21	1.55	-	5.51	5.51
3	6-15	1.412	64.3	3.12	4.44
4	6-15	1.412	64.3	3.93	5.25

Figure 9 also shows the apparent effectiveness of ferric sulfate in increasing filter yields at approximately the same cost as polymer only at this higher fly ash to sludge ratio. Line 2 is characteristic of a high dose fly ash. The steep slope shows the effect of fines sealing off the cake.

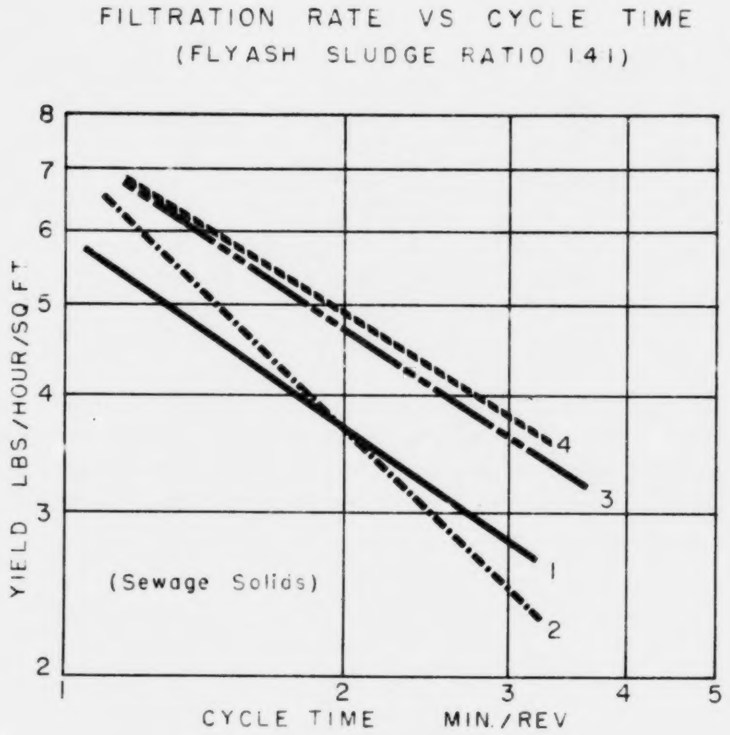


Figure 9

Sludge discharge at highest filtration rate (shorter cycle time) is apparent also. Normally the thinnest dischargeable cake at the shortest cycle time will give the highest yield, however, a point of diminishing returns can be reached at very fast cycle times, when the hourly yield will actually lower somewhat. Practically the fastest cycle time is one minute per revolution.

Sludge Filtration vs. Fly Ash Ratio

Figure 10 illustrates the sludge filtration rate as a function of the fly ash to sludge ratio. The points shown are filtration rates taken from their respective graphs of filtration rate vs. cycle time. The cycle time of 2.00 minutes was chosen because it is a medium cycle time and represents a point usually well defined on all filtration curves. The points used for this curve are as follows.

Line	Date	Yield	Fly Ash/Sludge Ratio	Cost per Ton
1	7- 5	3.80	1.36	$5.00
2	6-29	4.34	1.07	4.09
3	5-31	3.60	0.760	3.62
4	6-13	4.50	1.00	7.16
5	6-21	3.55	1.55	5.51
6	6-15	4.70	0.956	7.33
7	6-15	4.60	1.415	5.07
8	6-22	3.85	0.951	5.60

The optimum fly ash to sludge ratio appears to be approximately 1.0 to 1.1/1. At this point, the sludge filters well, and gives a good dischargeable cake of moderate moisture content, above this point (higher fly ash ratio) filtration of sludge may improve, but the economic advantages are diminished.

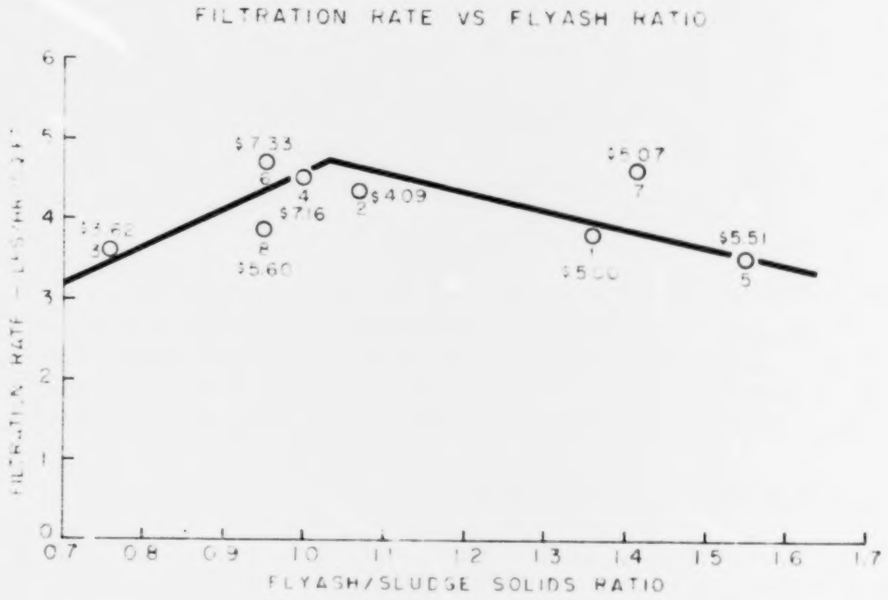


Figure 10

Cake Moisture

Moisture content in the filter cake is a function of chemical dosages and fly ash concentration. Figure 11 represents the range in moisture content encountered during the test runs. It can be readily noted that the addition of fly ash causes a significant decrease in moisture content of the filter cake.

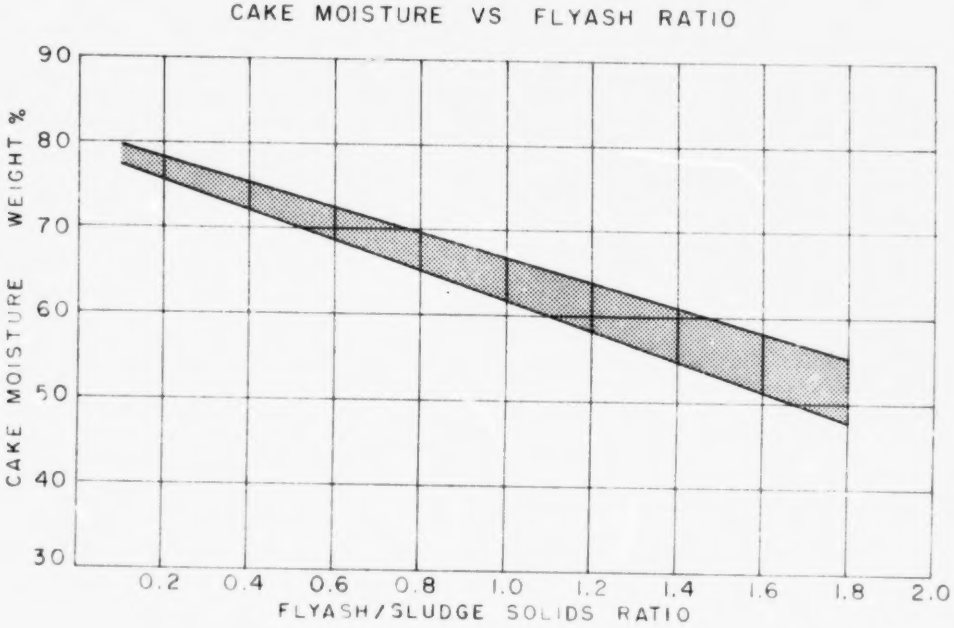


Figure 11

Conclusions

1. Fly ash is a valuable sludge conditioner. Preliminary studies indicate that its use lowered the total chemical costs of sludge conditioning from $20 to $4 per ton dry sludge solids.

Optimum fly ash to sludge ratio for dewatering performance lies between 0.9 and 1.4. However, the optimum economic fly ash to sludge ratio may be different due to costs of trucking and materials handling of ash and chemicals, and other operating costs.

2. The value of ferric sulfate is not clear-cut. It appears to be an effective flocculant at higher (1.4/1) fly ash to sludge ratio, but its benefit is not clearly defined at lower fly ash ratios. Additional investigation is warranted in a full scale plant.

3. It has become clear that an efficient vacuum filtration system must be under close control by the laboratory. Any automatic equipment which might be incorporated in final plans for the close surveillance of sludge characteristics should be evaluated.

SECTION V

PRESSURE FILTRATION OF SECONDARY DIGESTED SLUDGE WITH FLY ASH FILTER AID

Introduction

Prior to the pressure filter program the City of Cedar Rapids investigated various other methods of dewatering bottom sludge from their secondary digesters. A laboratory model pressure filter and a pilot scale pressure filter were operated for a period of five months using fly ash as a filter aid. The information collected during the initial study and presented in the following report represented nearly all of the analytical data available on the operation of the pressure filter in the United States.

In October 1967, a pressure filter with a filtration area of 36 square feet was installed with the necessary accessories at the Cedar Rapids Water Pollution Control Plant to conduct a pilot plant study for dewatering digested sludge. Fly ash from a local steam generation plant burning fossil fuel was used as a filter aid. Operation extended over a five month period with daily observations recorded. For a seven week period, a two shift operation was in effect to obtain more data in the limited time the equipment was available.

The pilot scale pressure filter may be generally described as a plate type, consisting of a series of plates about 24 inches in diameter, covered with a synthetic filter media. Sludge is center fed to the plates at an operating pressure of about 260 pounds per square inch (psi) for a period of one to two hours. At the end of this period, the plates are parted and the filter cake is discharged by gravity and the cycle is repeated. No conditioning chemicals were used, the only filter aid being waste fly ash.

At the time the pilot scale pressure filter study was undertaken, the pilot scale vacuum filter study had been completed. The results of the vacuum filter study have been presented earlier in this report. The purpose of the pressure filter study was to evaluate the pressure filter as a practical method of sludge dewatering in light of the earlier pilot plant study using a 3 foot diameter prototype vacuum filter. The vacuum filter study was considered successful and indicated that the Cedar Rapids digested sludge could be economically dewatered by that process. The basic objectives of the pressure filter study were as follows:

1. To determine the filterability of digested sludge at Cedar Rapids.
2. To determine the fly ash filter aid requirements.
3. To evaluate the pressure filter performance versus the vacuum filter performance reported earlier at Cedar Rapids.
4. To establish an estimate of cost for capital investment and operation of a full scale sludge filtration facility.

The pilot plant equipment was obtained from the Beloit-Passavant Corporation and was operated by personnel from the City Water Pollution Control Plant, and the Howard R. Green Company, Consulting Engineers, Cedar Rapids, Iowa.

Laboratory Model Pressure Filter

To determine the possibility of dewatering digested sludge using a pressure filter, a laboratory model filter was first used as shown in Figure 12.

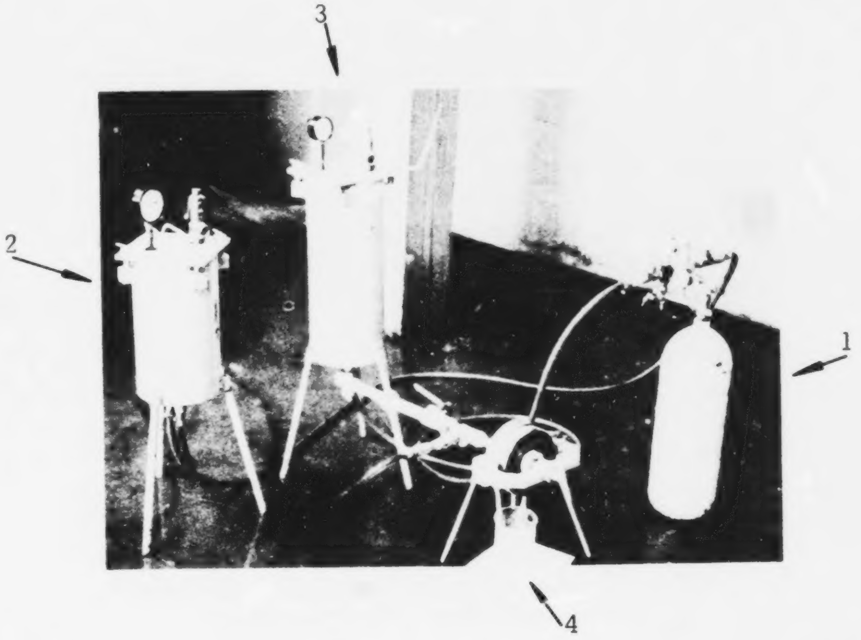


Figure 12

The unit consists of a pressure tank (1), which is a CO_2 gas cylinder. The gas cylinder, as a source of pressure for a laboratory scale experiment, is merely a convenience.

Unit 2, the precoat tank, is used to mix fly ash and water, and then to transfer it under pressure to the filter media at the beginning of the cycle. This mixture precoats the filter media for protection and improved dischargeability prior to applying sludge for filtration.

The feed tank, Unit 3, contains the sludge/fly ash mixture to be filtered. Immediately after precoat, this mixture is applied to the filter at a pressure of about 230 psi. The feed will continue for a predetermined period of one-half to two hours. The cycle time is of little concern, as the purpose of this experiment is to demonstrate whether a particular sludge is filterable.

The pressure filter, Unit 4, Figure 12 and Figure 13, consists of a split plate, about 6 inches in diameter, with an elliptical shaped void space, surrounded by filter cloth. Sludge is forced into this void under pressure of 230 psi and builds up on the filter cloth. Filtrate is removed in machined grooves behind the filter cloth.

Figure 13 shows the laboratory pressure filter in operation at Cedar Rapids.

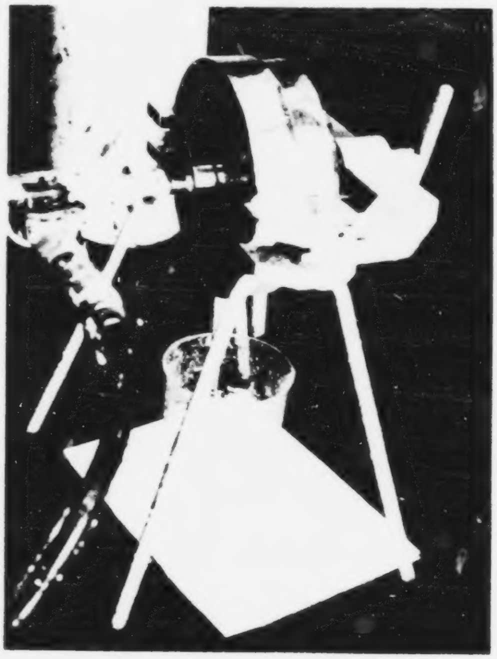


Figure 13
Laboratory Model Pressure Filter

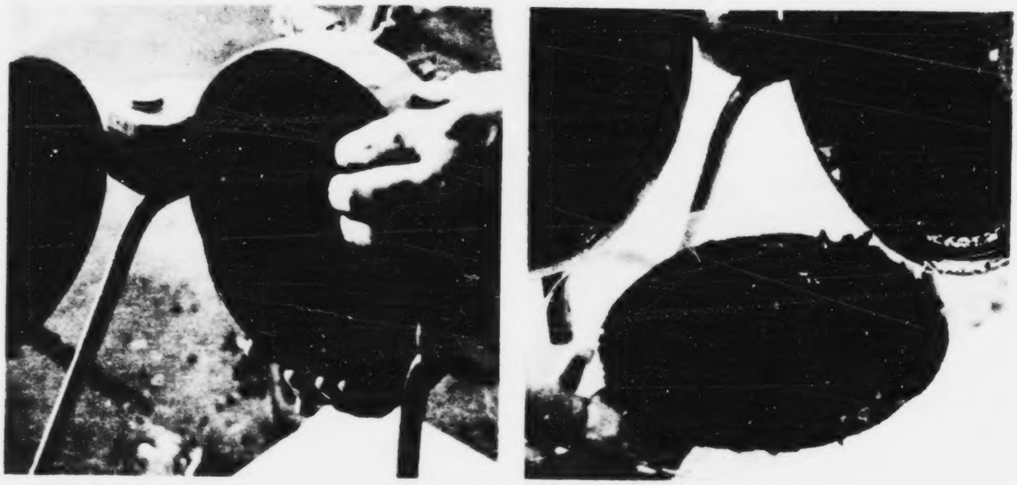


Figure 14
Removing Cake From Laboratory Model Filter

When the cycle is completed, pressures are reduced and the split plate is opened exposing the cake for removal as shown in Figure 14.

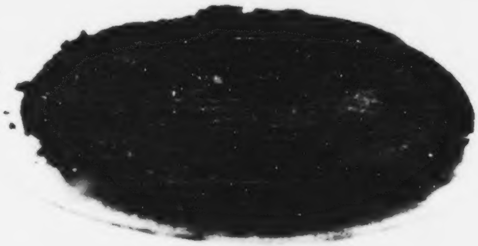
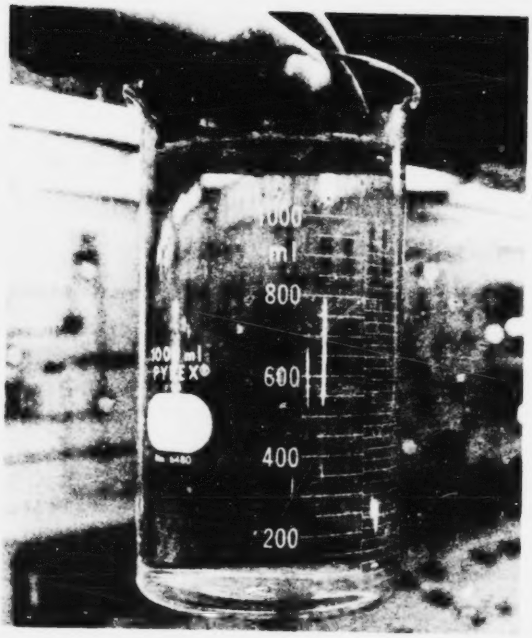
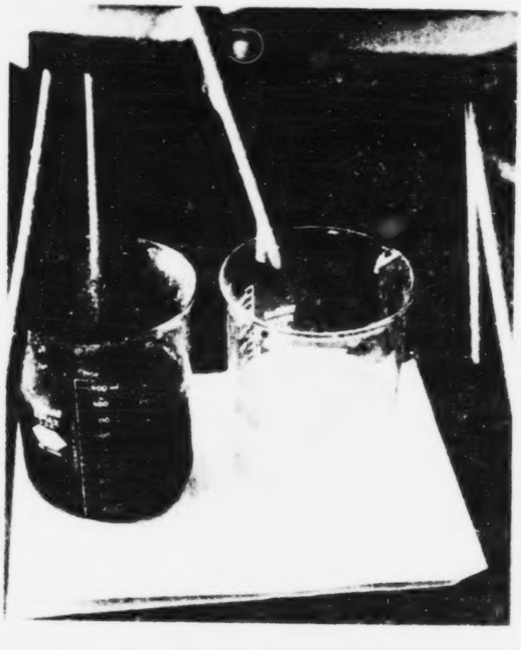


Figure 15
6" Diameter Filter Cake

A typical cake about 6 inches in diameter is shown in Figure 15. The center thickness is about 3/4 inches tapering to less near the edges.


1000
ml
800
600
400
200
PYREX

Note the Clarity of the Filtrate in the Beaker



Sludge Left Beaker - Filtrate Right Beaker

Figure 16

Pilot Model Pressure Filter

Description and Operation

At Cedar Rapids the source of sludge was from the secondary digesters. Sludge from these units was conditioned with fly ash obtained as a by-product from a fossil fuel power generation station.

The filter is a plate type, consisting of numerous plates in a vertical position, operating on a carrier. Each plate face is covered with a filter media, similar to materials used on a conventional vacuum filter. The filter media used for this study was a synthetic fabric, monofilament nylon. At the beginning of the filter cycle, all plates are closed and held as a pressure seal. Feed to the filter media is through the center of the plates.

Initially, the filter media is precoated with fly ash. The rate of application used at Cedar Rapids was 14 pounds wet fly ash per 36 square feet filter area. Ash moisture ran about 30 percent due to wet-down at the power generating station prior to open truck transportation and outdoor storage at the Water Pollution Control Plant. The dry rate of precoat was:

$$\frac{\text{14 pounds ash} \times \text{70\% solids}}{\text{36 square feet filter}} = \text{0.27 pounds ash/square foot}$$

This precoat is an arbitrary figure empirically determined on the basis of protecting the filter media and improving dischargeability. Precoat was applied at a pressure of 40 psi.

Sludge was placed directly from the digesters into a mixing tank and conditioned with fly ash. This tank contained a mechanical agitator and a sloped bottom which held the mixture in a homogeneous state during the filtering cycle. Fly ash in the desired quantity to obtain a given sludge/fly ash ratio was fed into this tank by a screw conveyor. This mixture is pumped to the filter feed tank to begin a filtering cycle. Air pressure applied to the filter feed tank transfers the mixture to the filter.

The sludge filtering cycle begins simultaneously with the ending of the precoat cycle and runs for a predetermined time or pressure. The mixture of digested sludge and fly ash was applied at an initial pressure of 100 psi and increased to 260 psi where this pressure is held until the sludge filtering cycle is terminated. The time lapse from start (100 psi) to beginning of maximum operating pressure (260 psi) was approximately 20 minutes (± 2 minutes). The filtering cycle was terminated at various time intervals, usually increments of one hour.

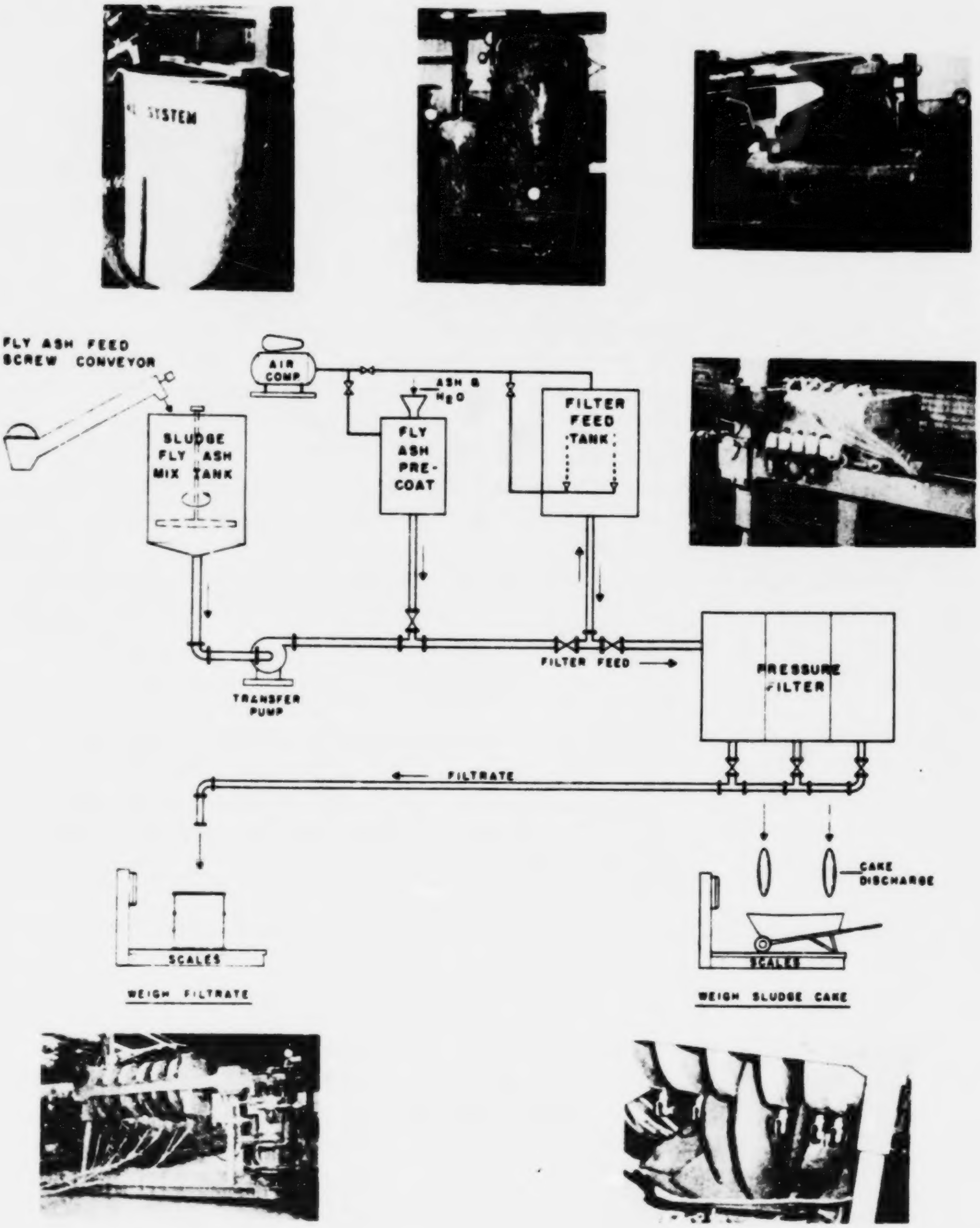
Filter pressure was developed by applying compressed air to the filter feed tank, which in turn, forced the sludge/fly ash mixture into the filter press at the established pressure. The same air source was used to maintain the filter tank mixture in a homogeneous state during the filtering cycle by means of agitation from a diffused air header on the bottom of the tank.

To complete a cycle, the sludge feed line is closed, and the wet sludge filter core is blown back to the storage tank. A reversing hydraulic operation of the filter plate carrier causes the filter plates to part, thus releasing the cake by gravity. The filter cake was caught in a wheelbarrow and weighed. In a full scale plant, the cake would be discharged to a conveyor. Following the removal of the sludge cake, the plates are closed, the filter is precoated, and the cycle is repeated.

A schematic flow diagram of the Cedar Rapids pilot plant is shown in Figure 17.

Pilot Filter Data

Filter	24" diameter nominal, carbon steel plates, gasketed with Teflon impregnated, braided asbestos. Dressed with polypropylene filter medium. (This can be changed to any other desired medium.)
Operating Pressure	225 to 260 psig (pounds square inch gauge)
Design Pressure	300 psig
Capacity	Six filter chambers Filtration Area: 36 square feet Cake Space: 2.25 cubic feet, 1½" thick cake
Closure	Hydraulic cylinder 7" x 30 strokes at 4500 psi
Hydraulic Power Pack	Operated by compressed air at 100 psig
Precoat Tank	60 gallon capacity - 22" diameter by 48" high, 200 psig water pressure


SYSTEM
FLY ASH FEED SCREW CONVEYOR
AIR COMP
ASH & H2O
SLUDGE FLY ASH MIX TANK
FLY ASH PRE-COAT
FILTER FEED TANK
TRANSFER PUMP
FILTER FEED
PRESSURE FILTER
FILTRATE
CAKE DISCHARGE
SCALES
WEIGH FILTRATE
SCALES
WEIGH SLUDGE CAKE

SCHEMATIC FLOW DIAGRAM OF PILOT PLANT PRESSURE FILTER

Figure 17

Feed Tank	300 gallon capacity - 36" diameter by 72" high, 300 psig water pressure
Sludge Mix Tank	48" diameter by 42" high open top tank polyester fiberglass with agitator, 30 RPM, 1 horsepower, gear motor
Air Compressor	2 stage, 2 cylinder, with 10 horsepower motor. Air receiver 120 gallon capacity for 300 psi water pressure.
Sludge Transfer Pump	Centrifugal, horizontal

Operational Experiences

No mechanical problems of significance occurred during the four months pilot plant testing period.

Considerable difficulty was encountered with the first two filter media due to extremely high grease content of the digested sludges. The high grease content of the digesters occurred during a period of internal reconstruction of the grease recovery system at a major meat processing plant. This problem was not immediately obvious until filter clogging began to occur. The early media was a yarn type, multifilament. Microscopic examination of the media, even after washing with grease cutting detergents, indicated that the grease particles could not be removed from the twists of the yarn configuration. The first media lasted forty cycles (80 hours). The second media of similar physical characteristics lasted twenty-six cycles (52 hours). When clogging occurred, it happened almost instantaneously.

> First Media - No. PP1Y1362, plain weave; multifilament polypropylene, thread count of 60 (warp) and 27 (woof), weight of 9.78 oz. per sq. yd., porosity of 8.5 cfm
>
> Second Media - No. PP2Y2362, Oxford weave, multifilament polypropylene, thread count of 72 (warp) and 22 (woof), weight of 11.8 oz. per sq. yd., porosity of 10-14 cfm

For the third media, it was decided to use a monofilament nylon which then performed satisfactorily the remainder of the testing period of 160 cycles, or in excess of 200 operating hours. At the end of the testing period, the monofilament media had no evidence of clogging or deterioration.

> Third Media - No. NY1S9611, plain weave; monofilament nylon, thread count of 107 (warp) and 101 (woof), weight of 3.38 oz. per sq. yd., porosity of 582.7 cfm

Sludge Characteristics

Sludge to be dewatered is bottom sludge from heated digesters and originates as thickened sludge from primary settling tanks, and secondary settling tanks following two-stage trickling filters (intermediate and final).

Sludge was removed directly from the digesters and conveyed to the pilot plant pressure filter site in a three compartment tank. Each compartment held approximately 150 gallons which was sufficient for one cycle, or three cycles per load. Sludge was used from all eight digesters, however, the majority of the sludge used for testing came from a secondary digester operated in series with a primary digester. In the future, all digesters will be operated in series as primary and secondary units, or stage digestion.

General Sludge Characteristics

Total Solids	4 - 7%
Volatile Solids	47 - 57%
pH	6.8 - 7.3
Volatile Acids	200 mg/l as Acetic Acid
Alkalinity	2800 - 3500 mg/l as $CaCO_3$

An extremely high ratio of industrial wastes to domestic wastes exists at Cedar Rapids. This is reflected in the range of raw sludge characteristics which often is also reflected in digester operation and its digester bottom sludge. This ratio of industrial to domestic wastes is in the range of 5 to 1 as measured in (BOD) population equivalent. In addition to domestic wastes from a population of approximately 110,000 persons, there are industrial wastes from meat processing; grain processing to starch, syrup, and cereal products; dairy operations; metal plating and over 100 miscellaneous smaller industries.

An example of extreme changing sludge characteristics during the pilot plant evaluation, which would not be normally anticipated, came from the meat processing industry. Municipal ordinances restrict the grease content of wastes discharged to the sewer to 200 mg/l. To comply with this ordinance, a major meat processing firm undertook extensive in-plant modifications, at which time considerable quantities of grease were by-passed to the waste treatment plant. This resulted in poor settling in the primary clarifiers and poor sludge thickener operation. The digester operation was frequently disrupted creating a low quality bottom sludge with high grease content. The early pressure filter runs were somewhat erratic because of this, and the filter media blinded due to high grease content. Blinding of the media was a very sudden occurrence, rather than a progressive one. However, some progressive blinding characteristics were observed in the precoat operation.

Filter Run No.	Grease Percent (% Total Dry Solids)
34	5.21
37	8.16
39	13.3
41	16.3
50	8.0
70	16.6
100	3.1
106	1.9
108	1.7
110	1.0

The grease content which caused so much trouble was reduced by improvements at the industrial source, and thereafter the filter operation was more uniform and dependable.

Fly Ash Characteristics

Fly ash used as sludge conditioner in this pilot plant study was obtained from a local fossil fuel fired, steam power generating station. At this plant, coal is pulverized and blown into the furnace for suspended combustion. The total ash is principally in the form of fly ash removed through cyclone collectors. This material is spherical in shape with about 95 percent passing a 200 mesh sieve.

At the present time, and during the pilot plant study, fly ash was transported in open dump type trucks from overhead bunkers. To move this material through the City, it was necessary to wet down the load. Consequently, the fly ash was received with varying moisture values. Outdoor storage also affected the moisture content. This varied from 23 to 35 percent moisture as it was applied to the sludge mixture. Fly ash, in dry condition as it is recovered at the power plant, has a tendency to "float" and is difficult to mix into the sludge mass.

For pilot plant operation, the moisture content was not critical because proper allowances could be made in the calculations, however, in a full scale plant, consideration should be given to receiving the ash in a dry condition.

Ash was hand fed through a screw conveyor to the sludge conditioning tank. The quantities were pre-calculated on a dry solids ratio after total solids determinations were made on both ash and sludge.

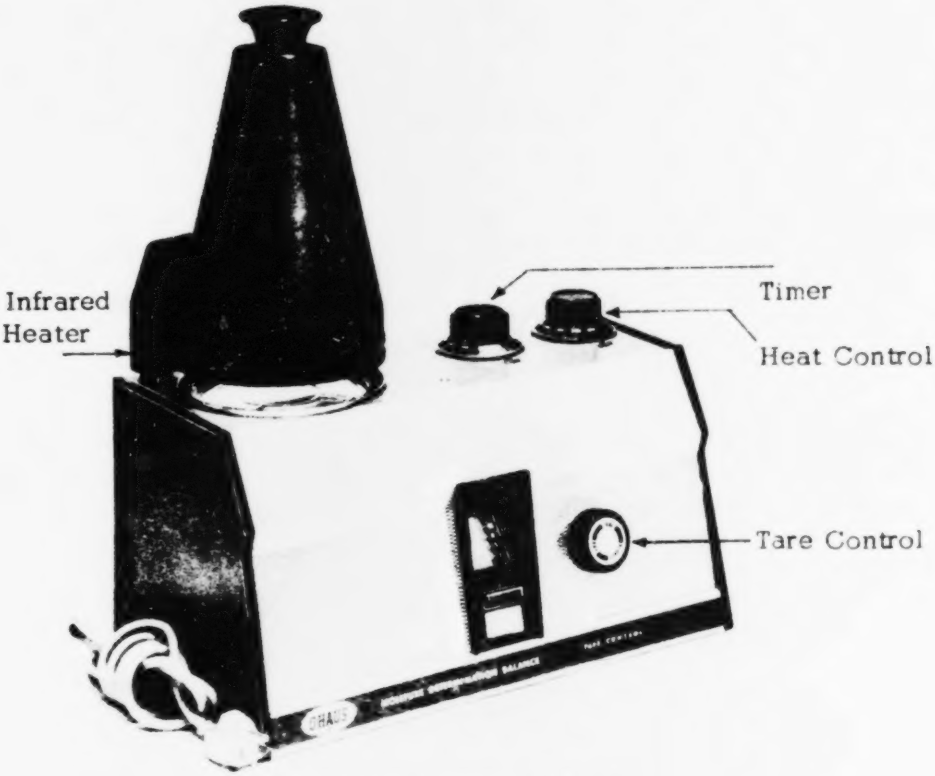
Automation of material handling facilities for a full scale plant would minimize pre-calculations and ash handling.

Laboratory Control

Primary consideration is given in laboratory control to the accurate and speedy evaluation of the dry solids content of the sludge, fly ash and cake. From this information most operational control data are determined. The principal control data are the fly ash to sludge ratio and specific sludge yield.

To accomplish the above evaluations it was necessary to divide the laboratory control program into two phases, on site and laboratory analysis. To determine sludge dry solids concentrations and fly ash moisture content on the site, an Ohaus moisture determination balance was used.

The moisture balance is a ten gram balance with optical projection read out to a hundredth (0.01) of a gram. Drying is done by an infrared lamp which has time and heat level controls to adjust for differing samples. Operation of the balance consists of measuring 10 grams of the sample in a tared aluminum dish and setting the heat and timer controls. The timer shuts off the heater after a predetermined length of time. When the sample has been dried, solids content is read directly to the nearest tenth of a percent (0.10%). Drying time is determined by plotting weight against time at various heat levels until a time and heat level is found which gives the fastest drying time without burning the sample. By using this balance it was possible to obtain the solids content of a sludge in 20 minutes and the solids content of fly ash in 5 minutes.



OHAUS MOISTURE BALANCE

The accuracy of the moisture balance was ± 4.0 percent of laboratory results. However, laboratory data are necessary to completely evaluate the facilities.

Two samples of the sludge were taken for each pilot run. These were composited in the laboratory and triplicate analyses for solids were performed. The sludge was also analyzed for grease content. Grease was determined by drying with magnesium sulfate monohydrate and extracting in a Soxhlet apparatus with petroleum ether.

A sample of fly ash and sludge admixture was taken and analyzed in the same manner as the sludge. This enabled back calculation of the fly ash and sludge contents. These samples confirmed the fly ash to sludge ratios obtained by dividing the dry fly ash weight by the weight of the dry solids in the sludge. This confirmed that there was adequate mixing in the sludge conditioning tank and that the sludge and fly ash samples were representative.

Two cake samples were taken by quartering the entire cake and completely mixing one quarter. The two samples of the mixed cake were then composited in the laboratory and analyzed for moisture in triplicate. All moisture samples were analyzed as residue by evaporation at 102° C. after drying 16 to 24 hours in a forced draft oven to constant weight.

Filtrate samples were grab samples taken during the run to determine the relative suspended solids concentrations. Suspended solids were determined by filtration through a glass fiber filter mat in a Gooch crucible. The suspended solids present in the filtrate appeared to be essentially precoat which was forced through the filter under the initial filtering pressure.

The filtrate was analyzed also for BOD and COD. The BOD samples were seeded to ensure an active aerobic bacterial population. Both the BOD and COD tests were performed according to the 12th Edition of Standard Methods. Good correlation was obtained between the BOD and COD, where the BOD equaled 0.6 COD.

Filtrate Characteristics

The filtrate flow rate was monitored by using a container and a stop watch. The filtrate quantity was then measured in a graduated cylinder and the flow rate computed.

Filtrate flow rate observation began at the end of one minute and continued to the end of the cycle. At one minute for the higher fly ash ratios (2.00 to 1), the flow rate is approximately 0.15 gallons per minute per square foot. At lower fly ash ratios of approximately 1.5, the filtrate rate is 0.10 to 0.12 gallons per minute per square foot. At the end of the cycle, the filtrate rate is about 0.007 gallons per minute per square foot. A typical filtrate flow rate versus time is given in Figure 18.

The filtrate from the pressure filter is nearly free of suspended solids. At the start of each cycle, the filtrate contains approximately 400 mg/l suspended solids but this rapidly diminishes to less than 10 mg/l suspended solids. It was determined by the physical characteristics and microscopic analysis that the small amount of suspended solids present primarily originated from the fly ash precoat and were not fines from the sludge itself.

FILTRATE RATE VS TIME

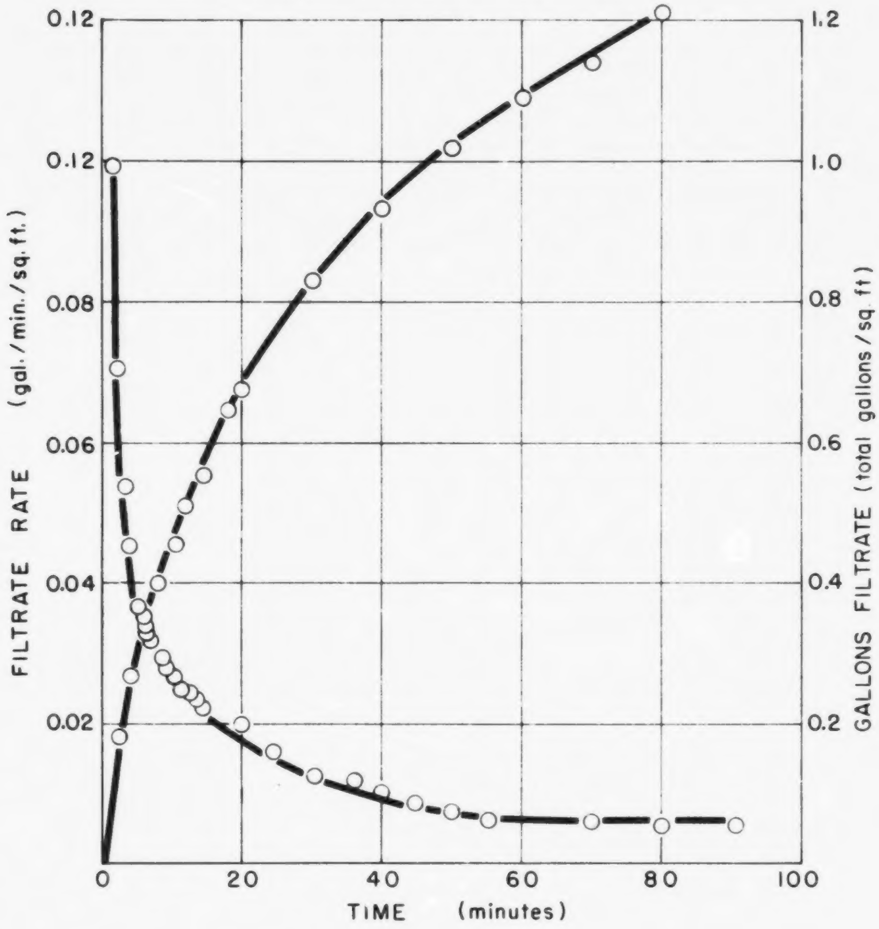


Figure 18

Figure 19 illustrates the suspended solids content at varying times during the filter run. Note the rapid decrease in suspended solids to nearly zero milligrams per liter. Because of the low suspended solids content of the filtrate, it can be discharged directly to the biological treatment unit.

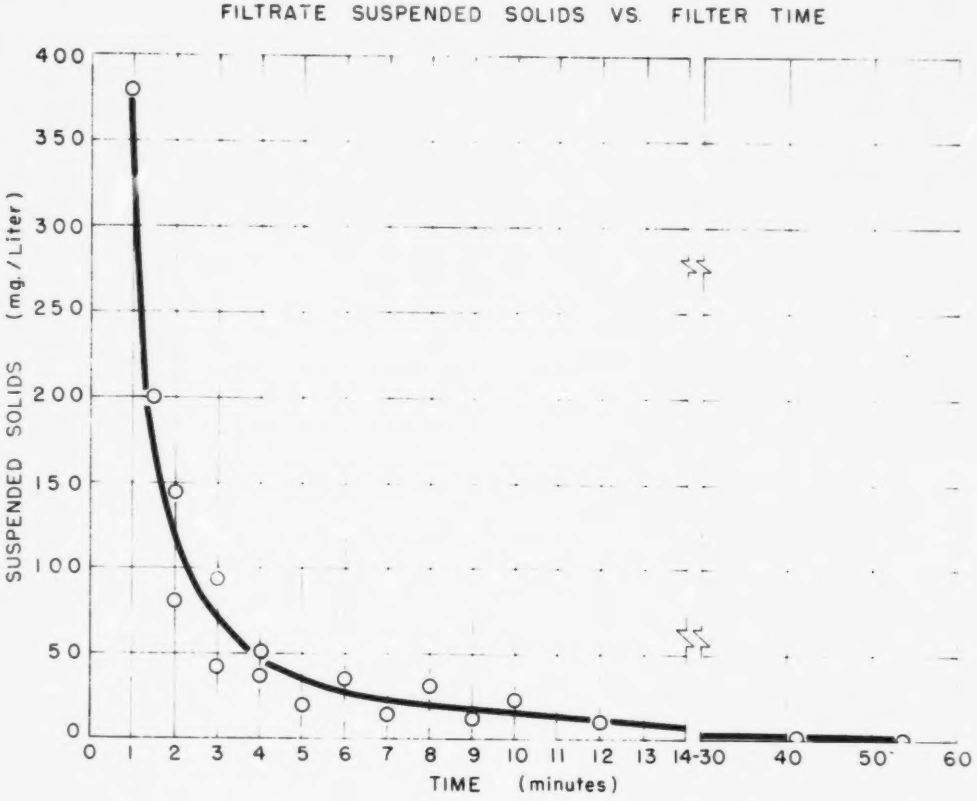


Figure 19

General Filtrate Analysis

pH	8.9 - 10.2
Suspended Solids	<10 to 400 mg/l
Total Solids	2000 to 3600 mg/l
BOD	180 to 1200 mg/l
COD	312 to 1700 mg/l
Color	Pale Yellow

The high pH of the filtrate results from the fly ash which contains several soluble alkaline metallic oxides such as calcium, magnesium, potassium, and sodium oxides.

Since the pressure filter acts as an efficient solids-liquid separator, the BOD in the filtrate is dissolved organic material. The BOD of the filtrate varies with the general conditions of the digesters. During the test period the BOD of the filtrate remained at the lower values of 180 to 300 mg/l for extended periods of time. At intervals the BOD of the filtrate increased upward to 1200 mg/l. This value in itself is not significant, however, the range of values is significant and probably reflects increased hydraulic and organic loadings to the digesters during periods of unusual industrial wastes loads to the treatment plant. The pounds of filtrate BOD to secondary treatment at the highest observation is such a small part of the total that it is insignificant to this study.

Test Results

The evaluations and conclusions in this report are based upon 222 complete pressure filter runs on Cedar Rapids digester sludge. Sludge was withdrawn from all digesters and every effort made to obtain representative samples for dewatering.

Considerable operating difficulties were experienced for a time due to unusual and varying characteristics of industrial wastes received at this plant. Particular problems occurred when wastes extremely high in animal fats adversely affected the digester operation and in turn reduced the filterability of the digester bottom sludge. Because of these problems some of the early filter data were erratic and did not classify well. This was frustrating to the operation of the pilot plant, however, it did reflect the potential problems characteristic of a plant operating with complex industrial wastes. These pressure filter operating problems resulting from sludge with high grease content were very similar to earlier ones experienced with the pilot vacuum filter at this plant. Fly ash used as a filter aid was obtained from a local power generation plant and was considered to be of uniform quality during the period of study. This was confirmed by analysis by the power company's laboratory at the beginning and end of the study.

Most work was performed by a two-man team. After the initial startup, it required about 20 minutes of cycle preparation, which consisted of precoating the filter, transferring the ash/sludge mixture, and charging the pressure filter. During the cycle, the second batch was prepared by one man, while the other man prepared laboratory data, or was free to perform assignments elsewhere.

The filter was large enough to render an approximate evaluation of the time required to operate a full scale facility. This equipment did not require constant operating attention. The total time spent at operation was dictated by the laboratory evaluations desired. It is assumed that a full scale plant would shut down automatically at the end of a predetermined cycle. The cake would be discharged and a new cycle started at the convenience of the operating personnel.

An operating data sheet was prepared for each filter run and was processed through the laboratory to include all of the desired data for that particular run. A copy of the data sheet is shown in Figure 20.

The initial step was to precoat the filter for protection of the media and to improve cake discharge. The precoat rate is an arbitrary figure and was determined to be 0.27 pounds ash per square foot as previously calculated in this report.

Filter runs were made over a range of digester solids varying from 3.2 to 7.6 percent, with the majority determinations in the range of 4.5 to 5.5 percent. It has been observed at Cedar Rapids that it is difficult to consistently produce digester bottom sludge above 5.5 percent. Sludge thickening should be considered in a full scale plant facility on digested sludge preparatory to dewatering.

Fly ash to sludge ratios are all expressed on the dry sludge solids basis. These ratios were varied over the range of sludge characteristics. This was done in an attempt to demonstrate the optimum fly ash/sludge ratio irrespective of material handling economics. The ratio was carried considerably above that which would be economical for municipal sludges.

Filter runs were classified into groups having similar characteristics of percent digester solids and fly ash/sludge ratios. These values were plotted to establish performance curves for Cedar Rapids' sludge and to ultimately develop design criteria to evaluate a full scale pressure filter sludge dewatering facility. This objective and procedure was similar to the earlier evaluation of sludge dewatering using a pilot plant vacuum filter.

The primary objective was to obtain a filtration rate (yield) in terms of pounds of dry sludge solids, per hour, per square foot of filter media.

Performance curves were established plotting the filtration rate (yield) in terms of pounds per hour, per square foot, versus time in hours for various sludge solids densities. Figures 21, 22 and 23 are for 4.5, 5.0 and 5.5 percent sludge solids.

A general observation indicates that as the fly ash/sludge ratio is increased to a given level, the filtration rate decreases. The optimum yield is the highest yield at the lowest fly ash/sludge ratio consistent with acceptable cake discharge and mechanical protection of the filter media. For this reason, a considerable amount of field interpretation is necessary and the plotted performance values are not an answer in themselves.

CEDAR RAPIDS PRESSURE FILTER PILOT PLANT

Date________ Run#________ Time Start________ Time End________

Digester #________ % Solids________ ________ ________ Avg.________

pH________ % Volatile________ ________ ________ Avg.________

Quantity of Sludge________ Gallons Vol. Acids________ mg/l

Remarks:________________________________

Fly Ash: % Solids________ ________ ________ Avg.________

Weight Wet________ x (%S________) = ________ Weight Dry

Fly Ash + Sludge: %Solids________ % Volatile________

Fly Ash/Sludge________ pH________

Precoat: Lbs. Fly Ash (Wet)________ x (%S________) = ________ Lbs. Dry

Filtering: Lbs. Pressure High________ to________ low. Length of run________ hrs.

Filtrate: Turbidity________ TS________ mg/l VS________ %

COD________ ppm. BOD________ ppm. Susp. Solids________ ppm.

pH________ Gallons________ Gal./SF/Hr.________

Filter Cake: % Solids________ % Volatile________

Weight Wet________ x (%S________) = Weight Dry

Dry Cake Density ÷ 2.25cf.________ lbs./cf. Total Gallons Filtered________

Yield (Lbs. Dry Cake ÷ 36 SF)________ lbs./SF Media

Filter Rate________ lbs./SF/Hr.

Remarks:________________________________

____

Figure 20

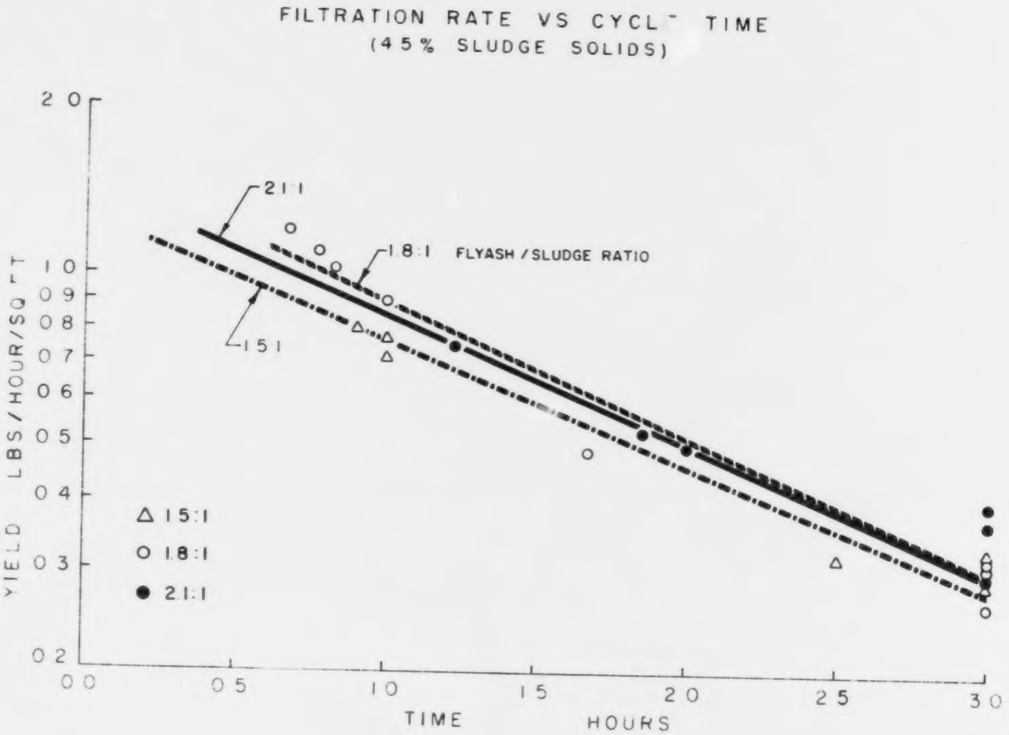


Figure 21

Cake discharge characteristics were observed at the end of each filtering cycle and evaluated to the fly ash/sludge ratio. As the filter plates are opened, the formed cake should break clean from the filter media and drop away as a single mass.

It was observed that satisfactory cake discharge characteristics could be maintained at fly ash/sludge ratios of 0.5/1 and with adequate precoat on the filter media satisfactory cake discharge should be maintained at much lower fly ash/sludge ratios.

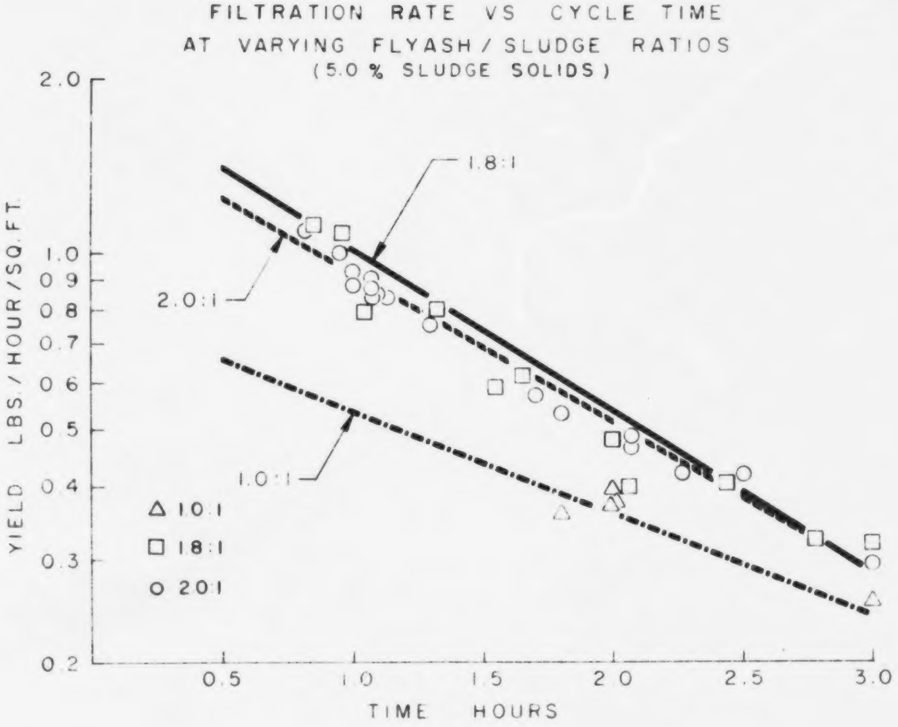


Figure 22

Plotting filtration rate versus fly ash/sludge ratio for a given percent sludge solids develops a reverse curve. As the fly ash/sludge ratio is increased the filtration rate is increased until an optimum ratio is reached. A further increase in the fly ash/sludge ratio causes a decrease in the filtration rate (yield).

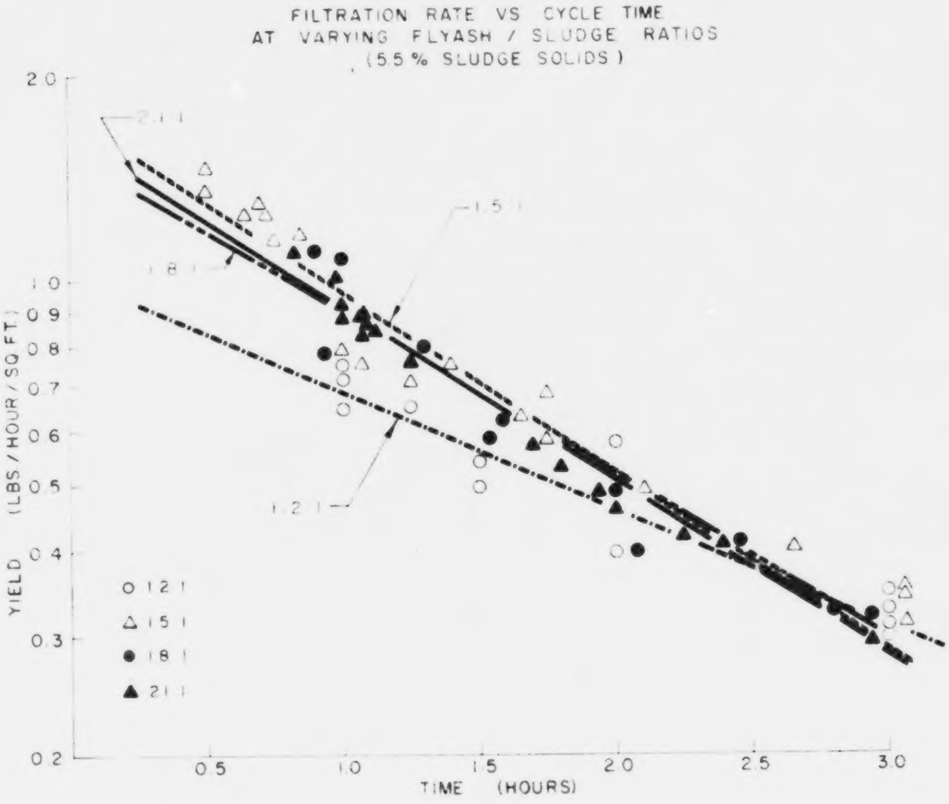


Figure 23

The optimum fly ash/sludge ratio appears to be less than 1.8/1, as shown in Figure 24 for 4.5 percent solids and Figure 25 for 5.5 percent solids for varying filter runs from 0.5 to 2.0 hours.

At the 0.5 hour (30 minutes) filter interval it was difficult to obtain a truly representative sample for analysis because the filter cavity was not completely filled with cake before beginning a new cycle. From an operational point of view the filter cavity should be completely filled with cake before beginning a new cycle and therefore, any filtering period of less than 0.5 hour (30 minutes) would not be practical for design purposes.

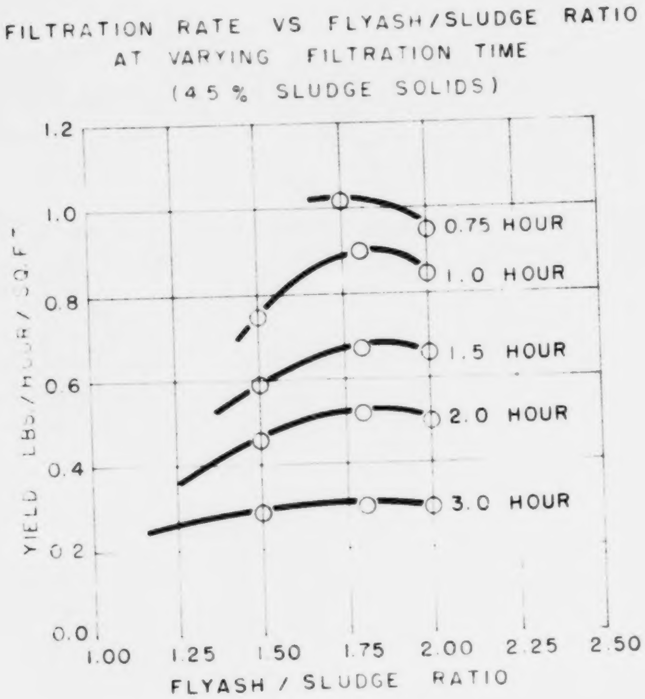


Figure 24

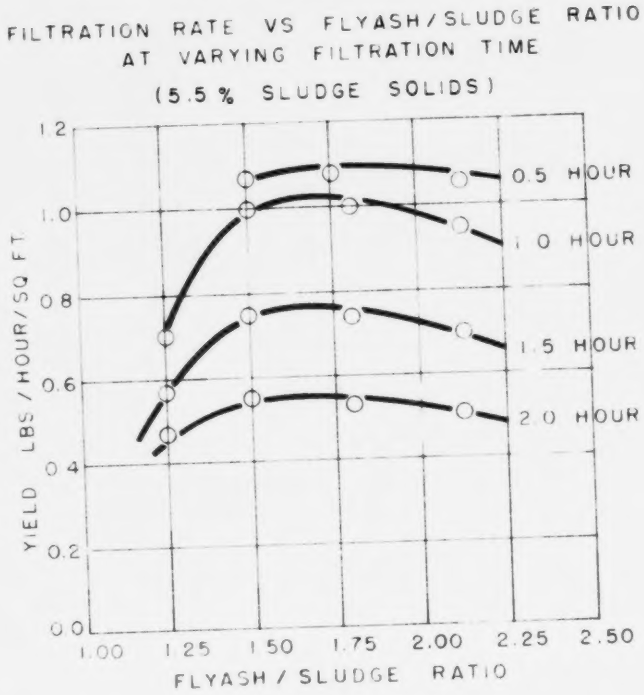


Figure 25

Percent cake moisture decreased at a rapid rate with an increase in filtration time up to approximately one hour. A very slight decrease in cake moisture values was experienced after about one hour of filtration.

This data is given as Percent Cake Moisture vs. Filtration Time in Figure 26 for 4.5 percent sludge and Figure 27 for 5.5 percent sludge. Note the similarity of curves for varying fly ash/sludge ratios.

The rate of development of cake quality is about the same for all fly ash/sludge ratios, however, the cake moisture value improves with an increase of fly ash/sludge ratio. Yield is expressed as percent of sludge solids only.

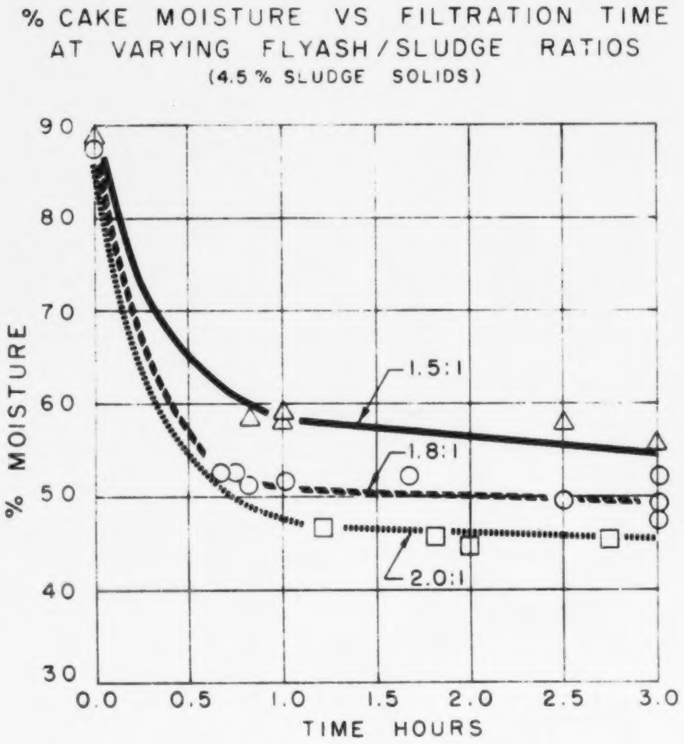
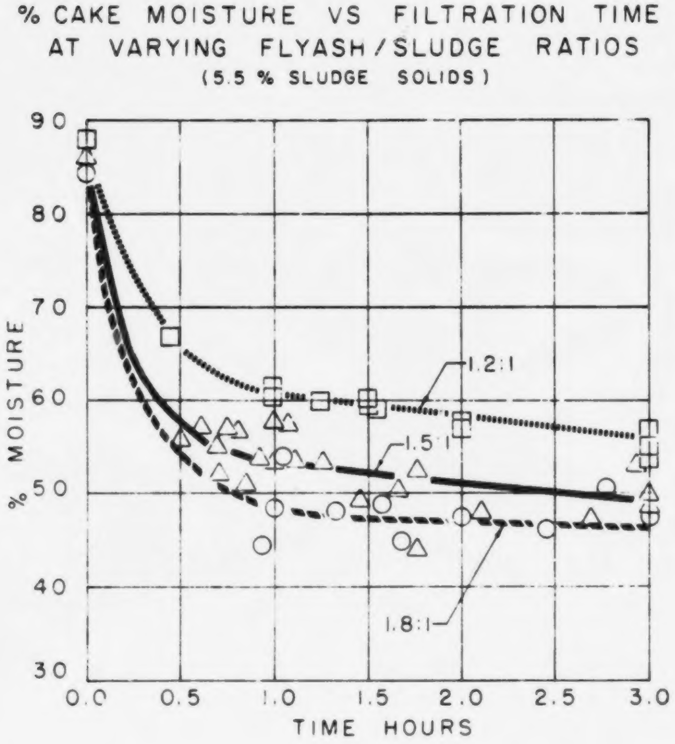


Figure 26



% CAKE MOISTURE VS FILTRATION TIME
AT VARYING FLYASH/SLUDGE RATIOS
(5.5 % SLUDGE SOLIDS)
% MOISTURE
90
80
70
60
50
40
30
0.0
0.5
1.0
1.5
2.0
2.5
3.0
TIME HOURS
1.2:1
1.5:1
1.8:1

Figure 27

SECTION VI

BASIS OF DESIGN FOR FULL SCALE PRESSURE FILTER DEWATERING FACILITIES

Introduction

Pilot test data compiled during the pilot scale pressure filter program was interpreted as a basis of design for a full scale pressure filter plant. During the period of study and of in-plant evaluations to plan solids handling facilities at Cedar Rapids, Iowa, numerous systems were evaluated and cost estimates prepared. The primary objective of these investigations was to determine the most economical system for dewatering sewage sludge at Cedar Rapids.

At the time of the original submittal to Research and Development for grant consideration in July 1967, pilot work had been completed using a vacuum filter and fly ash filter aid. The grant offer was made in March 1968 and during that time a pilot model pressure filter had become available and was operated at Cedar Rapids. Economic evaluations of handling power plant fly ash were less favorable than those of incinerated sludge ash and it was determined to install sludge incineration and recycle sludge ash as a filter aid.

For the application at Cedar Rapids the pressure filter and the multiple hearth incinerator was the most economical combination. Evaluating capital investment and full operating costs, the pressure filter-incinerator combination was estimated to be 10.3 percent less cost per ton of dry sewage solids dewatered and incinerated over the vacuum filter-incinerator combination. These cost comparisons are given later in this section. Other dewatering systems considered were even less favorable in cost comparison. Factors other than cost which influenced the selection of the pressure filter over the vacuum filter are also stated later in this section. Some of these favoring the pressure filter selection are: less building space, less operator attendance, closed system with fewer odor problems, drier cake, greater capture and clear filtrate, less power requirements, expandable capacity of filter unit.

The full scale pressure filter sludge dewatering facility was designed on the basis of experiences with the pilot scale pressure filter and with the attitude that the performance of the pressure filter could be improved upon. The equipment and the process was intended to be sufficiently flexible to allow a wide range of ash/sludge ratio to be evaluated,

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and also to perfect other approaches to sludge conditioning for improved dewatering such as chemical admixtures.

Any deviation from the intended program or process which may be conceived, developed or demonstrated, and which would improve the performance, should be considered and undertaken in both the plant design and in the plant study to follow construction. Modifications to the process, if successful, might accomplish one or more of the following:

1. Decrease operating time
2. Improve the quality of the product
3. Decrease the unit operating cost
4. Encourage the development of new methods and new processes
5. Give additional meaning to the Research and Development program

Filter Loadings

As discussed in Section V, Pilot Scale Pressure Filter, there appears to be an optimum fly ash/sludge ratio for maximum filter yield. As evidenced by Figures 21 and 22, this optimum ratio appears to be less than 1.8 parts fly ash to 1 part sludge at the lower sludge solids concentrations of 4.5-5.0 percent. However, at 5.5 percent sludge solids concentration, the optimum ratio of fly ash to sludge appears to be 1.5 to 1. (Figure 23). The apparent optimum ash/sludge ratio determined by the pressure filter performance may not be the optimum ash/sludge design ratio when evaluated with the total sludge dewatering and disposal process. Factors influencing the selection of the ash/sludge design ratio which are not equated in the pressure filter performance data include:

1. Material handling facilities
2. Volume of feed ash
3. Relative moisture in the cake product
4. Cycle time

5. Volume of ash/sludge product to be dried or incinerated

6. Unit heat requirements of mixture for drying

7. Unit heat value of mixture for incineration, and others

To provide continuity to the pilot plant study and this report, we sized the pressure filter and accessories on the basis of the performance at a fly ash/sludge ratio of 1.5 to 1, and reported out alternate performance at 1 to 1 fly ash/sludge ratio, and also 1 to 1 organic ash/sludge ratio.

Referring to either Figure 23 or Figure 25, at a fly ash/sludge ratio of 1.5 to 1, yield is found to be 0.75 pounds per hour per square foot of dry sludge solids at 1.5 hours' filtration time. Figure 27 indicates that the cake moisture would be 52 percent (cake solids 48 percent) which by normal sludge dewatering standards is a very dry cake. Further reference to Figure 27 shows that it would be impractical to extend the filtration time beyond 1.5 hours because the moisture content of the cake would not further diminish in proportion to the time used. On the other hand, the filtration time might well be decreased to about one hour and not appreciably increase the moisture content of the filter cake. For basic design, a 1.5 hour filtration time was used.

Between filtration cycles, during which time the cake will be discharged and preparations made for a new cycle of operation, an allowance of 30 minutes was made. This is a generous amount of time, however, it does not greatly affect the total cost of operation if it is estimated to be 10 minutes or 30 minutes. Unlike the vacuum filter, the pressure filter does not require constant attendance. It is, therefore, possible for the operator responsible for the pressure filter to be performing other duties between filter cake discharges, and he may elect to delay recycling immediately after the predetermined 1.5 hour filtering cycle. Or, the operator may quickly discharge cake and recycle in five or ten minutes and return to other duties to be delayed on the next cycle. The total hours of filtration would remain the same at 16 hours per 24 hour day.

Cake Production and Handling

It was planned to have the pressure filter and accessories as completely automated as was practical and to monitor most functions from a central console. Based upon operating experiences with the pilot scale plant, the method of controlling the filtration cycle would be:

1. At preset time interval of filtration. At the end of this time, filtration would cease and the operator would discharge the cake and recycle the process.

2. At preset pressure value. Theoretically, pressure should build up at an increasing rate as the filter cake is formed and could be used to determine the end of the filtration cycle when the filter cavity is filled. This theory of progressive pressure build-up might work in the filtration of a quality controlled manufactured product, however, it does not appear to be a useful control in the filtration of sewage sludges.

3. A preset rate of filtration. By this method the filtrate discharged from the pressure filter is metered and the filtration cycle is terminated at a given rate of filtration, such as 0.01 gallons per minute per square foot of filter media. A variation of this would be to meter the total volume of filtrate and the filtration cycle would terminate when a given value of total gallons was registered. This facility should be provided on all installations. Where it is not used to control filtration time, it may be used to monitor the physical condition of the filter media.

Time, pressure, and metered flow information will be desirable in the operation of the equipment and, therefore, all three methods might be readily available. This installation will plan to utilize the first and third methods. The time interval will be adjustable to changing conditions or requirements of sludge, ash, cake moisture and other admixtures. The rate of filtration will be metered to determine the end of the cycle. At the end of the filtration cycle, the operator would normally wish to discharge the cake promptly and recycle. However, with the pressure filter it will not be necessary for the operator to do this promptly if he is engaged in other activities. Unlike other filtration equipment, the pressure filter may be turned off at any time during the filtration process and left unattended for an indefinite period with no deleterious effect on operation or cleanup.

At the completion of filtration cycle, the operator standing at the control panel will prepare to discharge the cake. The procedure will be similar to that of the pilot filter. Sludge pumping to the filter will be terminated (if this was not done automatically as indicated above), the unfiltered sludge remaining in the filter core and feed lines will be discharged to the feed tank, conveyor equipment will be made to start, and the operator will proceed to the pressure filter.

Cake discharge is to be observed by the operator after activating a controller to open the filter plates. The cake will discharge freely and the large disc-shaped sludge cake will be sheared into small pieces by a series of steel cables strung across the receiving hopper. The broken filter cake will be transported by a conveyor system to the incinerator and then to ultimate disposal.

If additional sludge is to be dewatered, the operator will close the filter plates and recycle the process from an operating console. The sequence of filter precoat, sludge filtration, and cake discharge will then be repeated.

Filtrate Disposal

Disposal of pressure filter filtrate should present no problem. Filtrate from the pressure filter is nearly free of suspended solids and those present originate from the fly ash, rather than fines from the sludge solids. For a short period of time at the beginning of the filtration cycle filtrate rate may be relatively high. For the size of this facility, the filtrate rate may be about 650 gallons per minute for a few minutes. Within 10 minutes this value may decrease to about 25 percent of this higher rate and then continue to decrease to less than 10 percent of the maximum rate when the filtration cycle is half completed. Refer to Figure 19 for a typical filtrate rate of flow curve versus time. Filtrate may be handled in one of several ways.

1. Filtrate may be discharged to a relatively small holding tank having 10 to 30 minutes detention and then discharged at a lower rate directly to the biological (secondary) treatment units. The first 10 minutes of pressure filtration will produce approximately 33 percent of the total filtrate volume from the cake and at 20 minutes approximately 58 percent will be removed. (Figure 18.) Thus, with about a 7,000 gallon detention tank, the filtrate may be discharged at a rate of about 100 gallons per minute to the plant for secondary treatment.

2. Filtrate may be recycled back into the dewatering process as makeup water for the filter precoat with excess quantities discharged through the plant for secondary treatment.

3. Filtrate may be discharged to sludge thickening facilities where these are available and where the high pH of the filtrate might be an aid to sludge thickening.

Filtrate quantities should be metered on a continuous basis as an instrument of filter operating control. Filtrate rate of flow and total filtrate per time interval information can be used to adjust the filtration cycle and ash/sludge ratio.

Ash Requirements and Handling

Fly ash was available from two sources. Both sources are steam generating stations firing fossil fuel and are located approximately three miles from the treatment plant.

Fly ash should be hauled in closed trucks and blown into storage. Storage facilities should be provided for a minimum of 2 to 4 days' ash requirements. This will provide an adequate supply of ash to continue operation until other solids disposal arrangements can be made, should either ash delivery or cake disposal systems be temporarily out of service. Ash feed may be fed to the slurry mixing tank using volumetric feeders or controlled volume materials handling equipment. An adjustable speed screw conveyor may be calibrated with sufficient accuracy to handle the large volume of ash required. For this installation, gravimetric feeders were used to measure ash to the screw conveyor.

Solids Preconditioning and Thickening

Solids should be preconditioned in a sludge thickening facility prior to pressure filter feed conditioning. The pressure filter pilot study was conducted without sludge thickening because this facility was not available. However, sludge receiving facilities are necessary to control the application to the sludge dewatering equipment and these facilities may well serve a dual purpose. Any degree of sludge thickening will further improve the performance of the pressure filter. It was anticipated that the filter yield would be materially increased with thickened sludge.

Detention times in the digested sludge thickeners should be relatively long, in the range of 2 to 4 days' detention. With 5.5 percent solids entering the thickener, it might be possible to consistently come out with 8 to 10 percent solids. Small quantities of chemicals might further reduce the ash/sludge ratios required. It was planned to do research at this facility to determine the minimum ash/sludge ratio to obtain the maximum filter yield at the least cost.

Filter Cake Disposal

Cake discharged from the pressure filter having a moisture content of only 50 percent can be transported and disposed of as landfill or soil conditioner, providing it is pulverized for the latter use. Cake disposal in this condition would be limited to the City landfill, and sparsely populated areas for bulk dumping. Considerable quantities of pulverized cake might be used as a soil conditioner and fill material by the Parks

Department operating many municipal parks, golf courses and recreational areas. However, even at maximum utilization of this material by the Parks Department, a large volume will have to be disposed of by other programs.

To reduce the volume of material to be handled for disposal, and to provide an in-plant source of ash for conditioning, incineration facilities were installed. The proposed process will provide for disposal of solids at one or more points in the dewatering process to best serve the interest of both the process operator and ultimate disposal as follows:

1. Filter cake may be discharged in part or in total volume where soil conditioning might require maximum nutrient value or where landfill might require maximum bulk.

2. Filter cake may be dried in part or in total volume where the product might be stored for future use, or where wet filter cake might be offensive, and in this manner maintain maximum nutrient value and minimum bulk handling.

3. Filter cake may be incinerated in part or in total volume, thus minimizing the bulk handling requirements and utilizing the resultant organic ash as a sludge conditioner and filter aid. This will minimize ash handling requirements as well as solids handling requirements.

It was planned to both dry and incinerate by using a multiple hearth incinerator with provisions for reverse air flow when only drying is desired. This type of equipment affords the greatest flexibility of operation and with a reasonably low unit product cost. More detailed cost analysis appears later in this report. Modern facilities include gas cleaning devices to meet the local air pollution codes. The minimum operation and maintenance requirements of a multiple hearth incinerator are compatible with the minimum operating attention demanded of the pressure filter used as the primary equipment for sludge dewatering ahead of incineration. This combination will demand the minimum operator attention and may be operated by the same personnel.

Other systems evaluated and cost comparisons made for Cedar Rapids include:

1. Vacuum filter and rotary drier

2. Vacuum filter and flash drier

3. Vacuum filter and furnace incinerator

4. Vacuum filter and multiple hearth incinerator

5. Wet air oxidation and vacuum filter

6. Pressure filter and multiple hearth incinerator
    a. Fly ash filter aid
    b. Organic ash filter aid

Summary of Design Data

Digester Solids

Total Solids	56,000 lbs./day
Volatile Solids	26,400 lbs./day
Percent Solids in Sludge	5.5
Volume of Sludge Withdrawn	122,000 gals./day

Pressure Filter Loadings

	Fly Ash		Organic Ash
Ash/Sludge Ratio	1.5:1	1:1	1:1
Filtration Time per Cycle Hours	1.5	1.0	1.0
Total Cycle Time Hours	2.0	1.25	1.25
Total Filtration Time Hrs./Day	16	16.7	14.4
Cake Moisture Percent	52	60	50
Yield Filter Cake Lbs./Sq.Ft./Hr.	0.75	0.65	0.75
Filtrate Suspended Solids	Nil	Nil	Nil
Chemical Required	None	None	None

Other Plant Data

Population	1968 Present	113,000
	1988 Future	172,000
	BOD Equivalent	810,000
Plant Flow	Average, MGD	28.6
	Wet Weather, MGD	64.4

Estimate of Costs

The original estimates of cost of construction were based upon the August 1968 costs and were limited to the solids handling facilities directly related to sludge dewatering and incineration. For the convenience of this report, those 1968 costs have been scaled upward to August 1972 costs based upon the Engineering News-Record Cost Index of cost trends for construction and increased equipment costs. During the period since the original cost estimate (1968) and the present (1972), construction costs have increased about 54 percent and equipment costs about 20 percent. Other costs, such as operation, have inflated at least 4 percent per year.

The cost of operating the proposed pressure filter and incinerator was estimated from the power, fuel, materials and labor required at the design quantity and quality of sludge to be processed. These costs were determined with an electrical rate of 1.8 cents per kilowatt hour; off-peak natural gas rates of 4.8 cents per therm for the first 500 therms and 3.85 cents per therm over. Labor was determined on the basis of continuous attendance. Maintenance costs were based upon known cost of repair and replacement with consideration given to the equipment manufacturer's stated minimum estimates.

	Cost per Ton Dry Solids (Sewage)		
	Fly Ash		Organic Ash
Ash/Sludge Ratio	1.5:1	1:1	1:1
Hours of Operation	16	16.7	14.4
Power Pressure Filter	$ 0.60	$ 0.70	$ 0.54
Power Incinerator	0.48	0.40	0.18
Fuel Incinerator	1.46	1.68	1.70
Fuel Deodorize Stack Gases	0.98	1.06	0.24
Material Handling:			
Fly Ash to Plant	1.89*	1.26*	0
Incinerator Ash Disposal	1.25	0.97	0.34
Filter Cloth Replacement	0.15	0.16	0.14
Equipment Maintenance	0.52	0.56	0.56
Operating Labor	2.75	2.75	2.75
Total Estimated Operating Costs	$10.08	$ 9.54	$ 6.45

* The Power Company did not charge for the delivered fly ash, however, a minimum transportation charge likely would be charged in most cases. To compare processes on an equal basis an estimated charge is added for transportation.

Assuming amortization of bonds at 4¾ percent interest for 20 years, the capital cost per ton of dry sewage solids would be $20.00 based upon the estimated 1972 construction cost of $2,570,000. The total cost per ton dry sewage solids, including capital investment and operation, would be as follows:

	Fly Ash/Sludge Ratio		Organic Ash Ratio
Ash/Sludge Ratio	1.5:1	1:1	1:1
Capital Investment	$20.00	$20.00	$20.00
Operation & Maintenance	10.08	9.54	6.45
	$30.08	$29.54	$26.45

Valuations of Other Solids Handling Systems

During the period of planning solids handling facilities at Cedar Rapids, numerous systems were evaluated and cost estimates prepared. For the application at Cedar Rapids, the pressure filter and multiple hearth incinerator combination is the most economical and offers many additional advantages which are difficult to equate to a dollar value. Because of the great amount of extra work done in the plant operation, in the laboratory, and in design calculations, it is warranted to at least report briefly on the systems considered. It should be clearly understood that these evaluations are based upon sewage solids conditions at Cedar Rapids, and it is not intended to be a recommendation for other installations, nor does it imply criticism of those existing.

At the time of the original submittal to Research and Development for grant consideration, pilot plant work had been completed using a vacuum filter. This was in July 1967. The grant offer was not made until March 1968. During that time a pilot model pressure filter became available and was operated at Cedar Rapids and is the main subject of this report.

To effect a fair and equal basis for comparison of costs for several sludge dewatering systems, all previous cost estimates are projected to 1972 construction costs based upon the Engineering News-Record Index of Cost trends as previously stated. All sludge thickening facilities were omitted from these comparisons, and all other accessories were made to be exactly equal. As an example, all product haul was limited to three miles whether it be wet cake or incinerator ash. All costs of both the vacuum filter and the pressure filter were based upon performance at a 1:1 ash/sludge ratio because most of the vacuum filter work was conducted at this optimum ratio.

The estimated costs per ton of dry sewage solids were determined from the following sources.

1. All equipment costs are budget costs quoted by various manufacturers.

2. Building costs to house the equipment are estimated by the Engineer and are minimum building requirements.

3. Fuel and electrical costs are based upon equipment requirements and Cedar Rapids rates.

4. Labor costs are at plant rates.

5. Labor for operation and maintenance was determined by the Engineer and was considered approximately equal for all installations. This is conservative as it is reasonable to suspect that the vacuum filter maintenance costs are considerably greater.

Many advantages and disadvantages have been proposed by many authors and equipment representatives for each and every item of equipment. The merits of these are too often shaded by prejudice, and at best have little real value. Conclusions and observations from actual pilot plant performance using both a vacuum filter and a pressure filter at Cedar Rapids are shown on the following page.

			Operating Cost		
	Construction Costs	Amortize 20yrs.@$4\frac{3}{4}$% Per Ton	Dewater Only Per Ton	Dewater and Dry or Incin. Per Ton	Total Cost Per Ton
Vacuum Filter and Rotary Drier	$2,135,000	$16.55	$11.25	$25.70*	$42.25* Dry
Vacuum Filter & Flash Drying Furnace	2,155,000	16.75	13.05	23.00*	39.75* Dry
Vacuum Filter & Flash Drying Incinerator	2,600,000	20.40	13.05	20.05**	40.45**
Vacuum Filter & Multiple Hearth Furnace	2,080,000	16.15	11.25	16.80**	32.95**
Pressure Filter & Multiple Hearth Furnace (Fly Ash Aid)	2,570,000	20.00	5.00	9.54**	29.54**
Pressure Filter & Multiple Hearth Furnace (Organic Ash Aid)	2,570,000	20.00	4.12	6.45**	26.45**

* Dry cake approximately 12% moisture

** Incinerated ash

	Vacuum Filter	Pressure Filter
Near full time attendance required for one operation	Yes	No
Is a cleanup shift required?	Yes	No
Can one operator perform all shift duties?	Yes	Yes
Odors produced from operation	Considerable	Nil
Odor control accessories required due filter	Yes	No
Chemical required	Yes	No
Ash additive required	Yes	Yes
Cake, percent solids	33-38	40-55
Filtrate, suspended solids, mg/l	2000-6400	10-400
Cloth life, estimated hours	2000-4000	6000-8000
Power requirements, with accessories, KWH	1430	1080
Can filtration area be increased?	No	Yes
Operating Costs:		
Capital investment for equipment		20% more
Capital investment for building	50% more	
Incinerated costs per ton dry sewage solids	$32.95	$26.45

Obviously, there are many other items of comparison, some significant to treatment requirements, and many of little importance. The most important consideration, and the one which is often disguised by reporting it in part rather than in total, is how much will the total cost be for both capital investment and complete operation. As reported earlier, the total cost per ton of dry sewage solids disposed of strongly favors the pressure filter.

SECTION VII

FULL SCALE PRESSURE FILTER DEWATERING FACILITY

Treatment Plant Description

The full scale Solids Handling Facility at Cedar Rapids was planned in 1968, bids were received and contracts awarded in December 1969, and construction was completed in the spring of 1971. During the remainder of 1971 the sludge dewatering and incineration processes were checked out. The full scale Research and Development program for sludge dewatering with ash filter aid was carried out during the calendar year 1972, and is the subject of this report.

To better understand the size and operation features of the sludge dewatering facility, it would be well to briefly discuss the basic water pollution control plant from which the sewage sludges originate. An aerial view of the treatment plant is shown in Figure 28. The original plant construction was done in 1929 and at that time the plant served only that portion of the Cedar Rapids population served by sanitary sewers. Now the plant serves all of the metropolitan area, including Cedar Rapids, parts of the Town of Hiawatha, and numerous suburban residential areas. Primary and secondary biological treatment and separate sludge digestion was completed in 1933, with chemical-mechanical separate industrial waste treatment also constructed at that time. Expansion of all phases of the treatment processes, along with the addition of intermediate stage trickling filters was made in 1955. Separate sludge thickening and supernatant oxidation facilities were completed in 1965. A second major plant expansion, whereby the plant capacity was doubled to the present capacity, was completed in 1966-67.

The entire facility, except the final clarifiers, is covered with thin-shell domes and the exhaust air ozonated for atmospheric control. Over 8 ac. (378,500 sq.ft.) of process units are under cover to assure high efficiency during cold weather and to control the potential source of odors. Five separate air handling facilities evacuate the air from under the domes at a rate of over 120,000 cfm and deliver this air to a detention chamber after ozonation. The ozone installation represents the largest installation of its kind on waste water treatment with a capacity of 150 pounds per day. The largest domes are 167 feet in diameter.

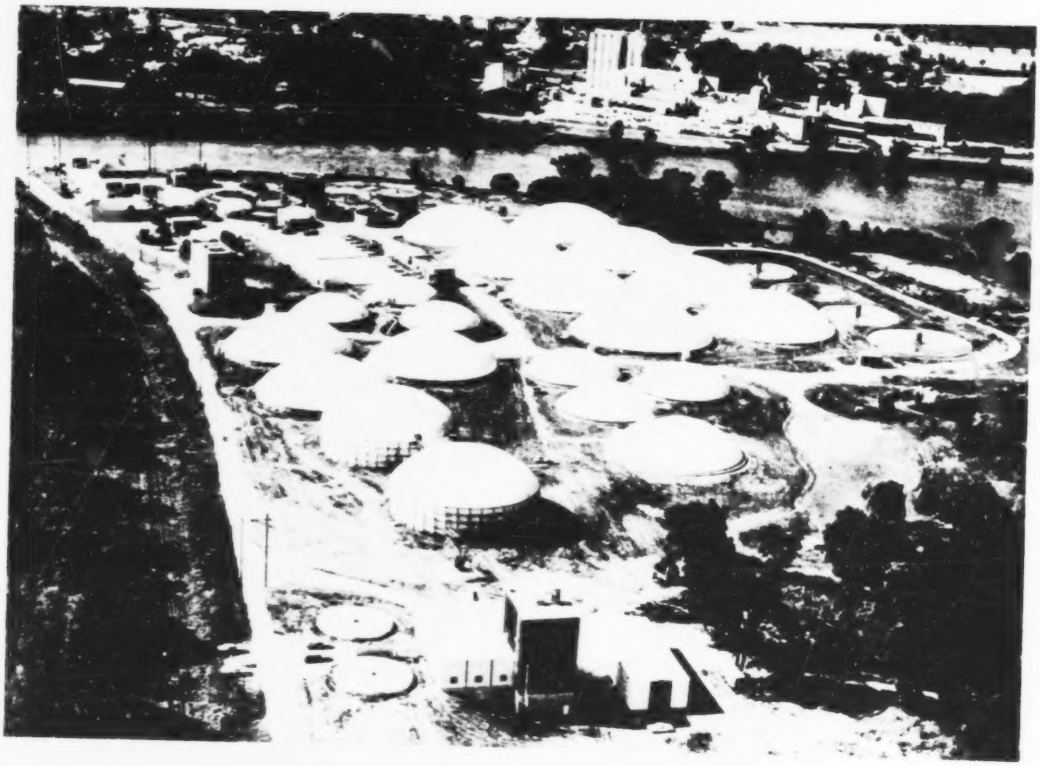


Figure 28

The plant processes are:

Primary	Screen chambers Raw lift pumping Grit removal Preaeration Primary clarification
Secondary	Plastic media roughing filters (first stage) Intermediate clarifiers Final rock filters (second stage) Final filters

Separate sludge digestion, raw sludge thickening, supernatant oxidation, digested sludge dewatering and incineration.

Flow	28.6 MGD average 64.4 MGD maximum

Organic	Raw BOD	135,000 pounds per day
	Effluent BOD	8,500 pounds per day
Population	Approximately	113,000 persons
	Equivalent BOD approximately 800,000 persons	

The digested sludge dewatering and incineration facility was completed in 1971. It consists of parallel trains of sludge dewatering facilities followed by multiple hearth sludge cake incineration. The two sludge dewatering trains consist of 50 foot diameter sludge holding tanks, sludge pumping and metering, ash feeders, sludge-ash mix tank, sludge-ash contact tank, conditioned sludge feed pumping, pressure filter, filtrate removal, cake storage bunker, followed by a multiple hearth sludge incinerator.

General Description of the Pressure Filter Process

Digested solids are periodically pumped from the secondary digesters to the solids holding tanks. Bottom solids from the sludge holding tanks are removed by pump on a demand basis and sent through preconditioning facilities to the pressure filters where the sludge is dewatered.

The sludge preconditioning facilities consist of sludge grinders; mix tanks where ash and/or chemicals are added as a filter aid; contact tanks where slow mix of filter aids and flocculation occur; and variable rate sludge pumps to feed the filter.

The pressure filters consist of a series of plates covered with a nylon, or similar, filter cloth. Sludge is pumped through the filter leaving a solids deposition on the filter cloth. This solids formation, or filter cake, is periodically discharged at the end of the filter cycle. The filter cycle may be 1 to 2 hours depending upon numerous variables.

Cake discharged from pressure filter is broken and then fed to the multiple hearth furnace for drying or complete incineration. Filtrate removed from the solids in the pressure filter is discharged back through the plant for further treatment.

The multiple hearth furnace may be operated as a cake drier, or as an incinerator. As a drier, the cake will be removed from the bottom and used as a soil conditioner at about 12 percent moisture. As an incinerator the combustible materials in the cake will be destroyed and the ash remaining will be disposed of as a low grade soil conditioner. The volume of incinerated ash will be a small fraction of the volume of dried cake. The complete process is observed from a panelboard, and operating records established. The basis of design is:

Total solids to digester-lbs/day	92,000
Total solids from digester-lbs/day	56,000
Volatile solids from digester-lbs/day	26,400
Percent solids in sludge	5.5
Gallons sludge withdrawn per day	122,000
Period operation-hrs/week	112
Ash to sludge ratio	1.5 to 1
Filtration time per cycle-hours	1.5
Cake moisture - percent	52
Yield filter cake-lbs/sq.ft./hr	0.75
Cake production with fly ash:	
Sludge-lbs/day	56,000
Fly ash-lbs/day	84,000
Total solids-lbs/day	140,000
Moisture 52%-lbs/day	152,000
Total cake-lbs/day	292,000
Operating schedule-hrs/week	112 (plus startup and shutdown)
Incinerator feed rate-lbs/hr	18,250

Pressure Filtration Technology

It is not the purpose of this report to expound upon the theory of filtration, or specifically pressure filtration. There are volumes of literature published on the theory and practice of sewage sludge filtration, most all of which relates to vacuum filtration, with an occasional comment on pressure filters or variable volume filters. In this country there is a very limited reference to pressure filtration theory and practice. This project was researched and developed by the use of pilot scale filters, starting with bench work and progressing to a small sized, production model pressure filter pilot plant as previously described. At that time there was no readily available analytical data in this country on the operation of the pressure filter. However, if there had been other detailed performance data available, the pilot plant approach for the development of the dewatering facility would still have been the recommended approach.

Pressure filter knowledge was not sufficient to provide information on which to predict or calculate performance. It is still a fact that the state of present knowledge is very limited and it is necessary to examine each material in the laboratory and, if possible on the pressure filter pilot plant, to determine its filtration characteristics, to obtain some indication of filter performance, and to provide basis to calculate a full scale plant design. A brief review of the pressure filtration process may be of interest. The object of any sludge dewatering process is to recover the solid content in a concentrated form suitable for disposal or further processing such as incineration, and to dispose of a filtrate relatively free from suspended solids. There are different kinds of cake filters, however, they can be conveniently grouped under four headings.

1. Vacuum filters, batch or continuous.
2. Pressure vessel filters, vessels operated under pressure containing elements upon which the cake is formed.
3. Pressure filters, filter presses, plate and frame, or recessed plate.
4. Variable volume filters, variable chamber, tube press.

The pressure filter which is the subject of this report is item three (3). It may have been described by numerous names, such as plate and frame press, or filter press. Literature has reported that about 1900, a number of plants in England installed sludge filter presses, principally those plants using chemically precipitation type of treatment processes. Plants were also reported at Providence, R.I. and Worcester, Mass. in the United States. These units were reported to consist of a number of cast iron recessed plates fitted in a frame between two head blocks which were forced together by screw jacks. The plates were 36" in diameter with corrugated faces for drainage channels and covered with filter cloth. Sludge was fed to these cloth covered recesses (chambers) through a 6" center opening by pump pressure or through an ejector using compressed air. Filtrate passing through the cloth was drained from individual spigots from each chamber.

Houston, Texas installed two 120 plate presses in 1919 to dewater activated sludge which operated at a maximum 100 pound pressure. Operation was discontinued after a short period of time due to cloth problems and high costs.

Literature indicates that the early pressure filter installations were not very successful due to premature blinding of the media, excessive leakage contributing to poor housekeeping, low capacities, high labor requirements to remove the cake, long filtration periods, and poorly fabricated plates which frequently failed. European interests stimulated in the late 1950's has lead to the present more efficient pressure filter. This has been followed in the past few years with interest in the United States for further development and improvements. An example of more recent improvements is the design to provide a separate plate gasket to assure longer media life and less leakage. The earlier pressure filters, and some of the present units, use a filter cloth sack which covered the entire plate and also served as the gasket on the joints of the plate. The filter media life was often controlled by the mechanical wear, or pressure wear as the plates were held together. Most often the cloth failed as a gasket long before it deteriorated as a filtering media. Separate plate gaskets control leakage if properly installed and adjusted and minimizes housekeeping. This gasket arrangement also greatly extends the useful life of the filter media. Details of a typical pressure filter plate are shown in a cross-section view in Figure 29.

There are a number of other forms of solids-liquid separation systems, some of which have a wide range of application, and some of which have often been used in domestic and industrial sludge dewatering. In the case of waste treatment sludges where the material contains a high proportion of fine particles, the separation of the solids-liquid phase is extremely difficult, and the force required to displace the liquid phase and replace it with sludge particles to produce a cake formation increases. In this application the use of applied pressure as in the use of the pressure filter to develop sufficient force to cause the separation of solids-liquid offers a distinct advantage. The proper sludge preconditioning, selection of an efficient filter media, controlled pressure development will improve the efficiency of sludge dewatering beyond that normally found in other systems. The product of pressure filtration is an increase in solids concentration in the cake with a decrease in suspended solids content of the filtrate. Compared to other systems for sludge dewatering the advantages for pressure filter may be summarized as follows.

1. The quantities of sludge conditioners are usually reduced. Conditioning materials may be of the least expensive variety (ash) as was the case at Cedar Rapids.

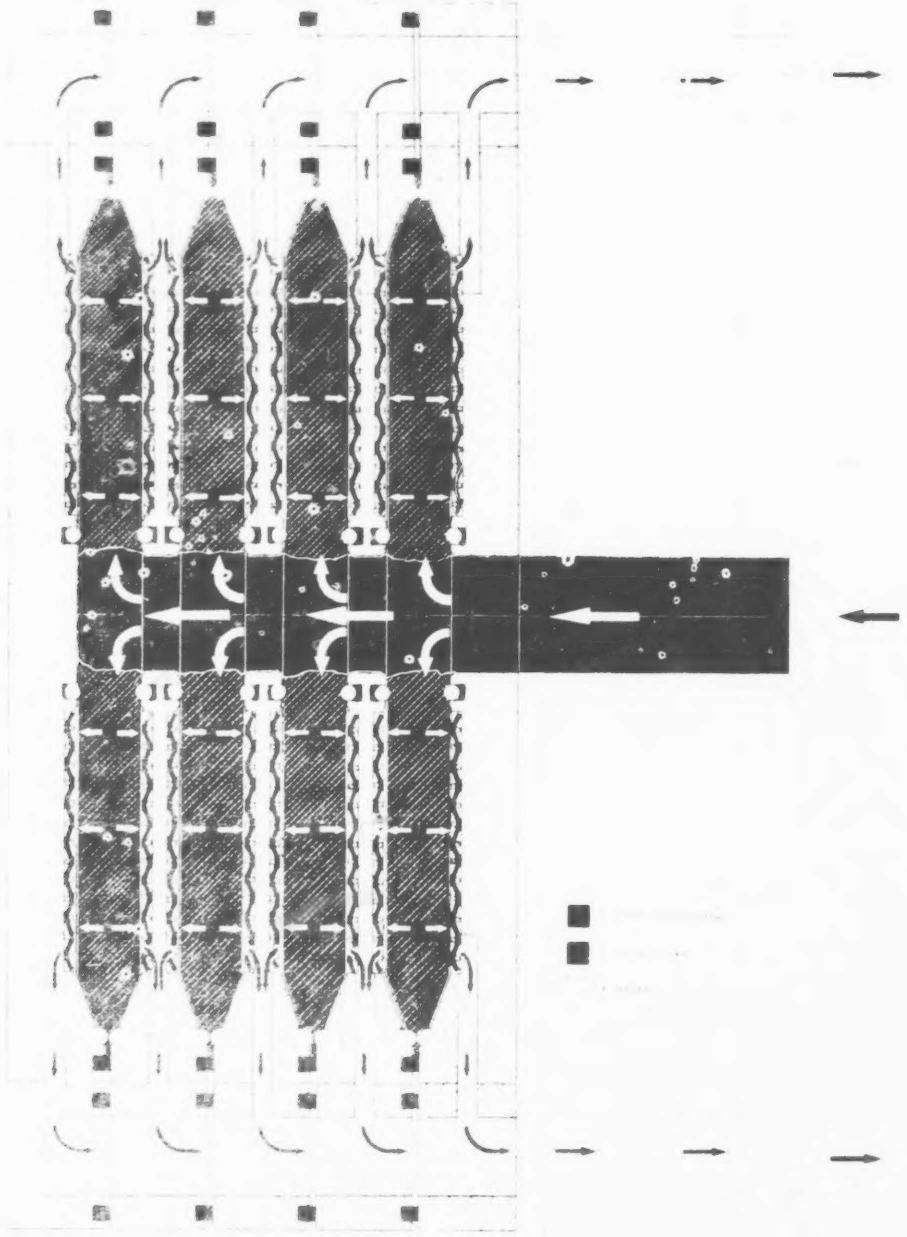


Figure 29

2. Filtration efficiency is maintained even when there is a substantial range of sludge characterictics, and is improved over most systems at normal filterability conditions.

3. Increased solids content of the filter cake. The low moisture content is important where cake is incinerated and often will assure autocombustion of the incinerator feed, thus saving supplemental fuel cost.

4. The filtrate is considerably lower in suspended solids and BOD, and thus minimizes the loadings back to the plant for treatment, or other problems of ultimate disposal.

5. Operation and manpower requirements are minimal, and considerably less than most systems. The attending operators may perform other duties during most of the period of filtration. This is not true of several other systems. Operators not skilled in mechanics can easily be trained to operate this filtration process.

6. Maintenance is minimal due to relatively few moving parts and standard supporting equipment such as pumps and mix mechanisms.

7. Filter cakes are more easily disposed of due to lower moisture content and smaller volumes. The cake may be conveniently stored in bunkers under the filter and removed by standard conveyor equipment and at continuous discharge rate thus overcoming any small effect of the batch process.

8. The costs based upon amortized capital costs are competitive with other processes. The initial capital costs may be higher than some other systems, however, the minimum maintenance costs extended over a significant period of useful life, and the low operating costs make the pressure filter system highly attractive.

The usual methods for evaluation of the filterability of a sewage sludge is the Buchner funnel test and the vacuum leaf test in the laboratory. The data is then applied to the standard filtration equations and established theory. This conventional laboratory investigation and filtration theory does not lend itself to pressure filtration because the majority of the pressure filter designs provided for a fixed volume cake formation. The cake is of a fixed thickness and occupies a fixed volume between the plates.

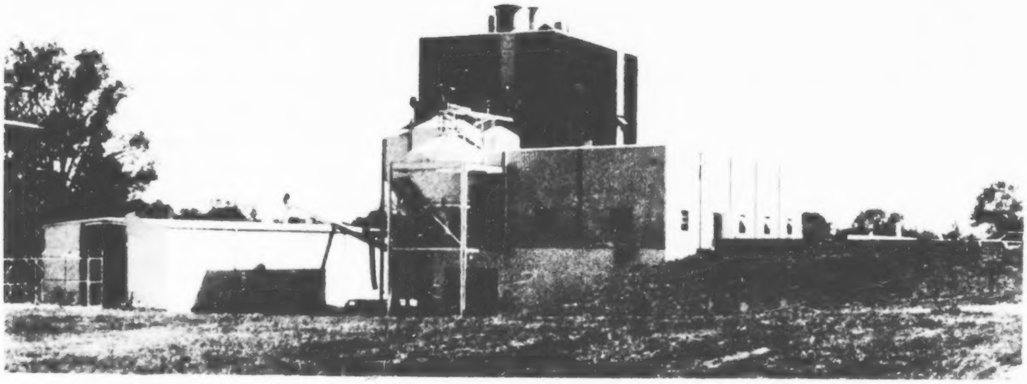
Cake formation also takes place under several varying conditions. At the beginning of the filtration cycle cake is formed at a high rate at relatively constant flow with the pressure increasing as the cake formation is developed. This is followed later in the cycle with a decreasing flow rate to a very small flow rate at a constant pressure drop. There are other variances from the standard filtration process which encourages the investigation using a pilot filter study to establish design criteria.

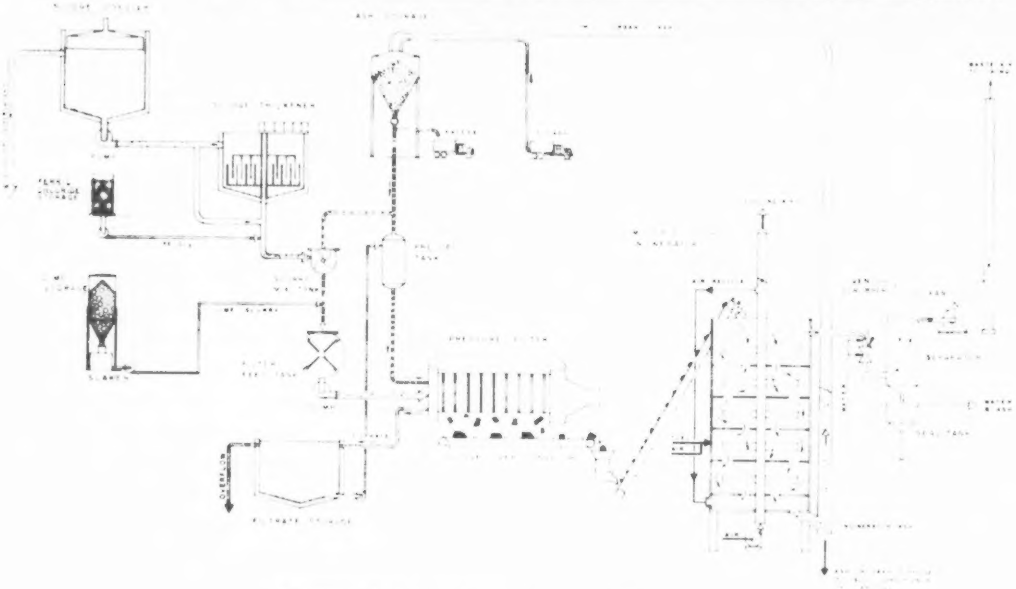
The present day pressure filter equipment offers many advantages and an efficient and economical system for dewatering sewage sludge. With a minimum of investigative study and pilot plant work, meaningful data can be obtained to size and evaluate a full scale pressure filter dewatering system.

Detailed Description of the Process

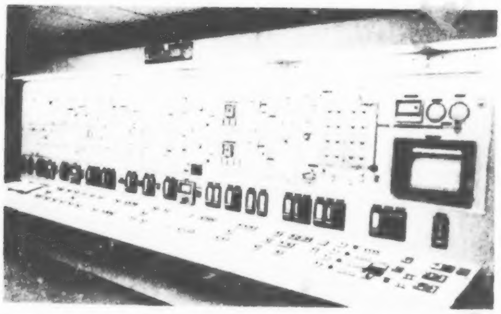
The process schematic and photographs of the facility are shown on Figure 30. Digested solids are removed daily from one or more of the secondary digesters and alternately transferred to one of the two 50 foot diameter sludge holding tanks. Each tank has about 2½ days' detention and sludge thickening mechanism. There was no previous experience that digested sludge thickening could be assured at this plant, however, bottom sludge removal mechanism was necessary so gravity sludge thickening with the picket design was installed in anticipation of some degree of sludge thickening. These holding tanks may be overflowed to a scum well for scum removal and may be decanted if thickening occurs. The scum may be preconditioned with chemicals and sent through dewatering, or it may be pumped back through the plant for further treatment.

Sludge from the bottom of the holding tank is pumped to the mix tank which is the beginning of the dewatering process. In the pump discharge line is a nuclear density gauge to measure and record the sludge density in terms of percent total solids. Also included in this line is a magnetic flow meter to register the volume of liquid flow pumped to the mix tank. With these two instruments a total accounting of the sludge pumped to the process can be accurately determined at all times. Both the nuclear density gauge and the magnetic flow meter are registered on the instrument panel in the operating room. Installed in the pump discharge line to the mix tank is a sludge grinder. The purpose of this grinder is to emulsify the sewage sludge solids to a uniform consistency and thus improve preconditioning and filterability of the sludge liquid applied to the pressure filter. The sludge grinders have replaceable cutting plates and blade assemblies and may be externally cleaned. Periodic cleaning may be necessary and may be detected by observing the pressure drop across

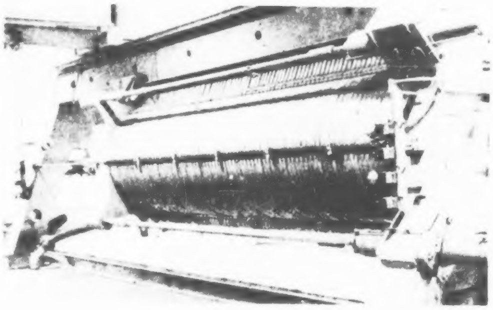



ASH STORAGE
PUMP
SLAKER
PRESSURE FILTER
OVERFLOW
FILTRATE STORAGE

SCHEMATIC SLUDGE DEWATERING & INCINERATION



Control Center



Pressure Filter

Figure 30

the grinders. Bypass valving and piping has been provided to bypass the nuclear density gauge, the sludge grinders and the magnetic flow meters, should the units be taken out of service for repair.

The mix tank is designed to blend the incoming sewage sludge with the filter aid which may consist of incinerated sewage sludge ash, fly ash from a local power plant, chemicals, or any combination of these. The mix tank is a horizontal steel tank, approximately 120 cu. ft. capacity, equipped with two speed agitators designed to thoroughly mix the sludge slurry and the ash and chemical admixture. Ash is fed to the mix tank by screw conveyor and is accurately measured by gravimetric feeders. Chemicals are added to the mix tank from a metered ferric chloride feed pump and from a lime slaker using a magnetic flow meter for control. The admixtures are controlled from the instrument control panel at the operators' room. Ash dust is controlled with spray showers in the mix tank and the operator may observe the slurry mixing process through a plexiglass observation door in the top of the tank.

Slurry overflow from the mix tank is gravity fed into the contact tank where it is further mixed at a very slow rate to provide ample contact between the sludge and the admixtures of ash and/or chemicals. This tank is also of horizontal steel construction with a slow speed agitator to maintain uniform consistency of the slurry. The tank has a large volume of about 600 cubic feet to assure an adequate supply of slurry feed to the pressure filter during each cycle of the filter operation. The contact tank is equipped with liquid level controls for starting and stopping sludge pumping and the mixing system. When the liquid level in the contact tank falls to a predetermined low level, the mix tank operation will start and a supply of slurry will be fed into the contact tank. As the pressure filter feed is satisfied, the liquid level in the contact tank will rise to a predetermined high level and shut the mix tank operation off. Slurry from the contact tank is pumped to the pressure filter by two filter feed pumps. These pumps are of a special design and are variable speed, variable capacity, variable pressure and are automatically controlled to decrease the output capacity as the pumping head increases until it reaches a maximum stall pressure of approximately 225 pounds per square inch, which is the normal high operating pressure of the pressure filter. This pump is hydraulically driven from a power pack which furnishes the hydraulic pressure necessary to stroke the piston.

The filter feed pump discharges to a surge tank to smooth the flow and to minimize pulsating pressures in the feed line to the pressure filter.

The pressure filter operates on a process cycle of about one to two hours duration depending on the density of the sewage sludge and the dewatering characteristics of the sludge slurry fed to the pressure filter. Prior to the beginning of the filter cycle the pressure filter is precoated to protect the filter cloth from blinding or clogging due to possible grease content or fine particles, and to form a shear plane between the cloth and cake to assure free and clean discharge of the cake. Filter precoat is a mixture of ash carried to the filter cloth in recycled filtrate water which is stored during each filter cycle for this purpose. Ash is charged into the precoat tank by a gravimetric feeder and conveyor system. All of these operations are sequenced automatically from the operator's control station by selecting the filtration cycle button.

Immediately after precoat the sludge feed to the pressure filter is automatically applied through motorized control valves to begin the filtration cycle, Sludge feed to the filter is from both the front and the rear ends and the rate of feed at the beginning of the cycle is very high and diminishes as the cake is formed. To accommodate this high initial rate of feed at the beginning of the filtration cycle, an equalization tank which was charged with conditioned sludge during the previous filtration cycle, and two filter feed pumps will discharge into the pressure filter. As the filtration cycle continues and the filter feed rate decreases, first the equalization will drop out of the system, and later one of the feed pumps will trip out of service. The second filter feed pump will continue to supply the feed to the filter at a decreasing rate until the filtration cycle has been completed.

The duration of the filtration cycle may be determined by time, pressure, or rate of filtrate flow. When the filtration cycle is determined by time, this may be one to three hours duration based upon previous filtration experience. It is not anticipated that the filtration cycle will be governed by pressure, however, should this be the choice, the filtration cycle would continue until a predetermined pressure level was developed, at which time filter feed would cease or would continue for a predetermined number of additional minutes.

Usual operation will be controlled by rate of filtrate wherein the filtration continues until a predetermined rate of filtrate flow is observed across a V-notch weir. An audio or visual signal is operated by a liquid level sensor at the weir.

At the end of the filtration cycle, the filter core feeding the individual filter plates remains filled with wet, soft sludge slurry. This soft core must be removed prior to opening the filter plates to discharge the cake.

This is accomplished by applying compressed air to the core at one end of the filter thus blowing the wet sludge core back to the contact tank through an energy dissipating cyclone. The energy dissipating cyclone is called a core expansion tank which discharges the wet slurry to the contact tank and vents the air to the atmosphere. At the completion of the core blow operation, the filter plates may be opened at any time convenient to the operator. Core blow and opening the filter plates for cake discharge is controlled from the hydraulic control console located near the filter. When the filter is opened each plate is progressively moved by a mechanical plate shifter located at the top of the filter and which continues to operate automatically unless stopped by the operator. The filter cake is discharged to the bunker below the pressure filter and is sheared into smaller particles as it passes shear cables stretched across the opening below the filter.

It is desirable that the operator be in attendance during the time the filter cake is being discharged. The operator should be certain that the cake has dropped free of the filter plates and should observe the condition of the filter cloth media, the plate gaskets, and other physical features of the filter. Should any large cake fragments hang up on the filter media the operator should remove these using a wooden paddle or similar approved device. The total time required to discharge the filter cake and recycle the process is in the range of 5 to 10 minutes. This is the time the operator spends at the filter.

Filter cake is discharged to a cake bunker under the filter and is conveyed to the multiple hearth incinerator by a series of cake conveyors and elevators. The cake bunker conveyor has variable speed control which serves as a cake feed rate control for the incinerator. All other conveyors moving cake to the incinerator are constant speed. If the incinerator is not in operation and it is still desirable to dewater sludge, an alternate discharge is provided to a truck for conveyance of wet sludge cake to a suitable disposal area. Conveyors are controlled from the operating console and are interlocked for proper sequence of operation.

The incinerator is a multiple hearth type, having nine hearths of 25 ft., 10 inch O.D. Temperature is maintained with automatic temperature controls by self combustion of the sludge cake and by auxiliary fuel fired burners using either digester or natural gas. Incinerated ash is discharged through a grinder to elevated ash storage hoppers. Combustion gases are removed through a high-energy wet Venturi scrubber operating at about 40 in. W.C. The system also has a deodorizing afterburner system. The incinerator is totally automatic and monitored at the operator's control panel.

Two (2) pressure filters are installed, each having 83 chambers of approximately 2.5 cubic feet volume per chamber, for a total of 207 cubic feet volume. The filters have 3400 square feet of filter area each and are equipped with an expanded frame for a future increase to approximately 100 chambers. The filter plates are carbon steel construction, epoxy coated. The filter media is a monofilament polypropylene with a stainless steel mesh backup screen and a carbon steel drainage screen. The filter operates at approximately 225 psig filtration pressure. The plates, during operation, are held closed by hydraulic cylinders operating with an oil pressure at approximately 5,000 psig. The pressure filter plates are gasketed with asbestos Teflon gasket material and the filter media is caulked in place with cotton caulking cord.

SECTION VIII

EVALUATION METHODS

Sampling

There were five basic sampling points for the evaluation of a pressure filter:

1. Sludge feed
2. Ash feed
3. Sludge-ash mixture
4. Filter cake
5. Filter filtrate

Sludge Feed

A sludge feed grab sample was collected by the operator from the discharge of the transfer pumps twice per operating cycle. The operator determined the sludge solids content of each sample using an Ohaus moisture balance in order to adjust the ash feed. The remainder of the sample was composited equally with the other sample for that cycle, refrigerated, and sent to the laboratory for analysis.

Ash Feed

Ash feed was analyzed on an infrequent basis. Samples were obtained from the ash feeders. A device was fabricated to collect the ash from a predetermined area of the gravimetric feeder belt. This ash sample was weighed and the quantity checked against the calibrated feed rate for the feeder belt. The device cut an ash sample one foot long in size and functioned on the principle of an ordinary cookie cutter.

Sludge-Ash Mixture

Sludge-ash mixture was sampled at the intake of the filter feed pumps. These samples were handled in the same manner as the sludge feed. Solids analysis on these by the operator confirmed the ash feeder settings. (ash/sludge ratio).

$$\text{Ash/Sludge Ratio} = \frac{\%\ \text{Sludge Mix}}{\%\ \text{Sludge Feed}} - 1$$

Example:

Sludge = 5%   Ash + Sludge = 17%

$$\text{Ash/Sludge Ratio} = \frac{17\%}{5\%} - 1 = 3.4 - 1 = 2.4$$

These results were later confirmed by the laboratory.

Filter Cake

The filter cake was sampled by grabbing representative pieces of cake when the cake was discharged and broken on the shear cables at the top of the cake storage bunker. These samples were sealed in glass jars.

Filtrate

Filtrate samples were collected from the weir box adjacent to each filter. Three grab samples of filtrate were taken during each cycle, composited and refrigerated for laboratory analysis.

Plant Measurements

All important measurements, such as pressures, ash weight, sludge volume, sludge density, time, etc. were monitored continuously by appropriate instrumentation. All measurements were recorded and integrated in the operators' control room.

Sludge Flow

Sludge flow was monitored by integrating magnetic flow meters.

Sludge Density

The percent of total solids of the sludge feed was monitored by nuclear density meters. The output (0-16 percent solids) was displayed on the chart as the sludge flow.

Lime Feed

Gallons of lime feed were measured by integrating magnetic flow meters. The concentration of lime was maintained at 0.4 pound CaO per gallon of water by appropriate adjustment of feed water rotameters and lime feed transmission at the slaker.

Therefore, pounds of lime feed equals gallons (from integration) times 0.4 pounds per gallon.

Ferric Chloride

Ferric chloride feed was by positive displacement feed pumps with calibrated discharge integrated from time of operation.

A 40 percent ferric chloride solution was used. Therefore, pounds of ferric chloride equals gallons solution times 4.75 pounds ferric chloride per gallon.

Ash Feed (Precoat and Admixture)

Ash quantities were metered by use of gravimetric belt feeder equipped with totalizers.

Filtrate

Total filtrate from the plant was measured by a 60° V-notch weir on the discharge of the filtrate storage tank. The depth over the crest was monitored by a bubble tube sensor.

Instrument Maintenance

All instrumentation was routinely calibrated by a qualified instrument technician.

Gravimetric feeders were calibrated monthly (minimum) by either weighing a timed discharge (Lime Feeders) or by removing material from a measured area of the weigh belt and checking against theoretical discharge (Ash Feeders).

Laboratory Analysis

The following routine laboratory analysis was performed.

Sludge Feed Analysis	Method
Total Solids	Dried at 104° C. overnight
Volatile Solids	Ignition of dried residue at 550° C.

pH	Glass electrode
Grease	Dried using magnesium sulfate extracted with hexane in a Soxhlet extractor
Volatile Acids	Determined by column chromotography according to the 13th Edition of Standard Methods
Alkalinity	Titration of the sample to pH of 4.5 using 0.5N H_2SO_4
Ammonia	Analyzed by distillation into boric acid and titration with standard acid
Organic Nitrogen	Kjeldahl method using mercuric oxide as catalyst
Sludge + Ash (Mixture)	
Total Solids	Same as above
Volatile Solids	Same as above
pH	Same as above
Filter Cake	
Total Solids	Same as above
Volatile Solids	Same as above
Nitrogen	Same as above
Filtrate	
pH	Same as above
Total Solids	Same as above
Volatile Solids	Same as above
Turbidity	Determined using a Hach Model 2100A turbidimeter

BOD_5	Dilution method (three dilutions per sample) initial and final dissolved oxygen concentrations measured with a Weston and Stack Model 350 dissolved oxygen meter with stirrer
Total Alkalinity	Titrated with 0.03N H_2SO_4 to a pH of 4.5
Color	Determined using a Hach color comparitor
Conductivity	Measured using a Labline Lectromho conductivity bridge with a cell constant of 1.0
COD	Measured basically the same as Standard Methods and except samples were autoclaved for two hours at a pressure of 15 psig and titrated with 0.25N $FeSO_4$
Total Phosphate	Determined according to Standard Methods. Potassium persulfate was used as the oxidant. Stannous chloride and ammonium molbdate were used for color development

Determination of Specific Resistance

The filterability of a sludge can be defined as the ease at which the sludge gives up water. While it is possible to talk about the filterability of sludge using subjective terms such as difficult, average, etc., it is desirable to be capable of objectively describing a numerical value to a sludge to give us a meaningful value for operation.

One method of evaluating a sludge is by determining its specific resistance. Specific resistance can be viewed as the reciprocal of filterability where a high value of resistance means a poor value for filtration and vice versa. The theory and development of specific resistance and its relationship to pressure dewatering have been developed by Dr. Oswald Busse of the Passavant Werk, Germany. This background material may be found in the dissertation "Entwicklung Eines Wirtschaftlichen Verfahrens Für Die Beseitigung Von Abwasserschlämmen" by Dr. Oswald Busse.

Specific resistance has proved valuable at Cedar Rapids in day to day operation. By using the specific resistance an operator can:

1. Determine whether a sludge is conditioned properly

2. Evaluate and optimize new methods of conditioning such as organic polymers

3. Evaluate the filterability of the unconditioned sludge

All these evaluations can be accomplished using a bench scale filter.

Specific resistance is measured by using a small pressure filter which is generally referred to as a resistance meter.

Mechanical Description of the Resistance Meter

The resistance meter is essentially a pressurized Buchner funnel. The filter is composed of a stainless steel body and support screen for a standardized filter paper that serves as the filter media. Meter accessories required are a 250 psig compressed air supply, a 100 milliliter buret and stand, stop watch, and convenient glassware as indicated in the photograph, Figure 31. A functional diagram of the resistance meter is shown in Figure 32.

Procedure for Using the Resistance Meter

A standardized, or calibrated, filter paper is necessary to assure a uniform filter media. This is wetted and placed in the meter on the support screen. A 100 milliliter sample of the sludge to be tested is measured and placed in the resistance meter. The top of the unit is secured and 250 psig pressure is applied. The effluent from the meter then drains into the 100 milliliter buret, which was initially filled to the lower 100 ml mark before the test. As the filtrate is collected in the buret, readings of filtrate quantity and time are noted and recorded.

In the case of an unconditioned sludge which has a high resistance to filtration the total filtrate is read after two minutes and at one minute intervals thereafter for a period of twenty minutes, or until the filtration has ceased and air blow-by occurs with air passing through the filter.

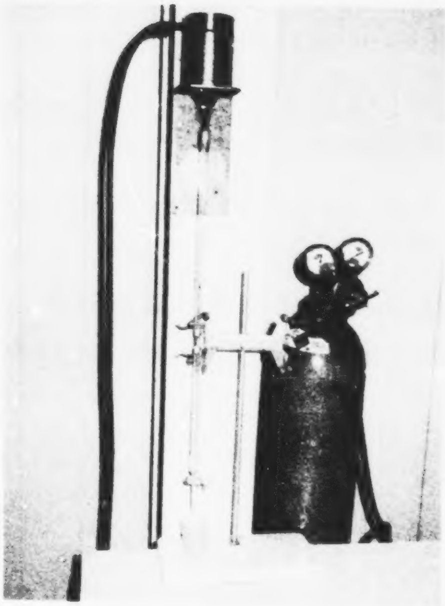
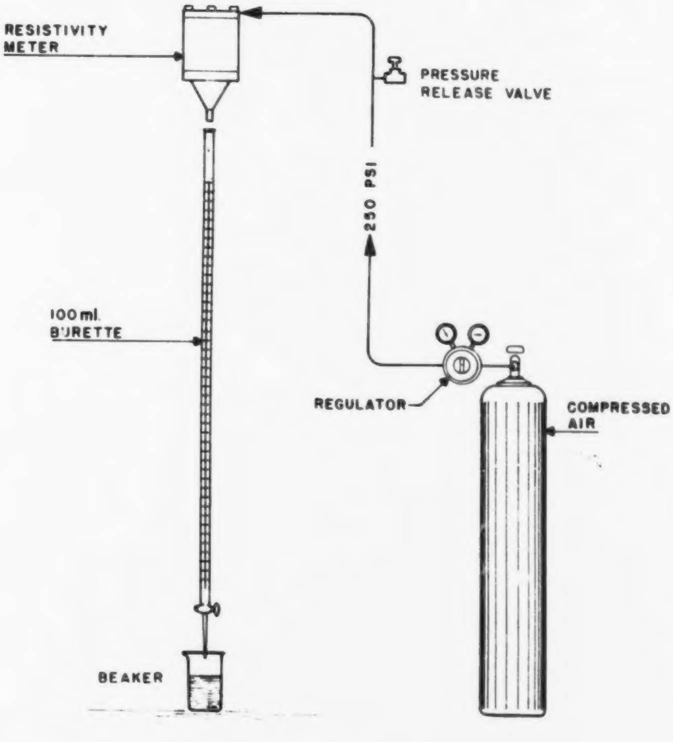


Figure 31


RESISTIVITY METER
PRESSURE RELEASE VALVE
250 PSI
100 ml. BURETTE
REGULATOR
COMPRESSED AIR
BEAKER

RESISTIVITY METER SCHEMATIC

Figure 32

In testing a condtioned sludge which yields water quite readily if the filtrate volume reaches the 75 ml mark (25 ml filtrate) within one minute, filtrate volume readings are then taken every 10 seconds. These readings are taken until blow-by occurs or an elapsed time of 200 seconds passes.

Following are two examples of resistance meter tests, one with an unconditioned sludge and another with a conditioned sludge.

Unconditioned Sludge pH = 7.4 % Solids = 6.0 T = 28.0°C.

Filtration Time Minutes	Filtration Time Seconds	Volume Milliliters	Seconds/ Milliliter
0	0	0	
1	60	1.3	46.2
2	120	2.3	52.2
3	180	3.3	54.5
4	240	4.1	58.5
5	300	4.9	61.2
6	360	5.6	64.3
7	420	6.2	67.7
8	480	6.8	70.6
9	540	7.4	73.0
10	600	8.0	75.0
11	660	8.5	77.6
12	720	9.0	80.0
13	780	9.5	82.1
14	840	10.0	84.0
15	900	10.4	86.5
16	960	10.9	88.1
17	1020	11.4	89.5
18	1080	11.8	91.5
19	1140	12.2	93.4
20	1200	12.5	96.0

Conditioned Sludge pH = 9.5 A/S = 1.5 CaO = 0.5%
Ferric Chloride = 0.5% % Solids = 6.0%
T = 29.0°C.

Filtration Time Seconds	Volume Milliliters	Seconds/ Milliliter
0	0	
10	8.6	1.2
20	16.5	1.2
30	22.6	1.3
40	27.7	1.4
50	32.4	1.5
60	36.6	1.6
70	40.4	1.7
80	Blow-by	

Analysis of Data

The equation for specific filter resistance is

$$r = 2 \times \frac{p \times Fi^2 \times b}{n \times c}$$

Where r = specific resistance in cm^{-2}
p = differential pressure^2 in dynes/cm^2
Fi = filtration area in cm^2
b = slope of line obtained in test by plotting seconds/milliliter versus milliliters
n = filtrate viscosity in poises
c = grams of dry solids/grams liquid for 1 gram sludge

In practice the above equation reduces to

$$r = k \times b/c$$

Where $k = \frac{2 \times p \times Fi^2}{n}$

When the pressure and filtration area are held constant, k then becomes a function variable only with temperature which governs the filtrate viscosity.

Therefore the constant k can be obtained from a table of viscosities of water versus temperature.

The remaining factors b and c are obtained by determining the density of sludge and the slope of the line obtained in the test by plotting seconds per cc versus cc.

In the case of an unconditioned sludge, Figure 33 shows a graphical solution of the slope b of 4.4. For this sludge the percent solids was 6.0, $c = 0.06/0.94 = 0.064$, temperature = 28.0° C. for a k of 5.315, therefore, r equals $k \times b/c = 5.315 \times 4.4/0.064 = 366 \times 10^{12} \text{cm}^{-2}$.

Also in the case of the conditioned sludge as shown in Figure 33, $b = 0.020$, k again is 5.315 and $c = 0.064$. Therefore, $r = 5.315 \times 0.020/0.064 = 1.66 \times 10^{12} \text{cm}^{-2}$.

Interpretation of Specific Resistance

In the two previous examples the specific resistance of an unconditioned sludge $366 \times 10^{12} \text{cm}^{-2}$, and a conditioned sludge $1.66 \times 10^{12} \text{cm}^{-2}$ have been determined. This data becomes meaningful when it becomes known what value of specific resistance a sludge must have to filter well. Generally speaking, at Cedar Rapids a conditioned sludge must have a specific resistance of less than $20 \times 10^{12} \text{cm}^{-2}$ and preferably less than $10 \times 10^{12} \text{cm}^{-2}$ to filter well. Knowing this limit we can optimize chemical or ash dosages using the bench model pressure filter much more easily and rapidly than a trial and error method on the full scale filter.

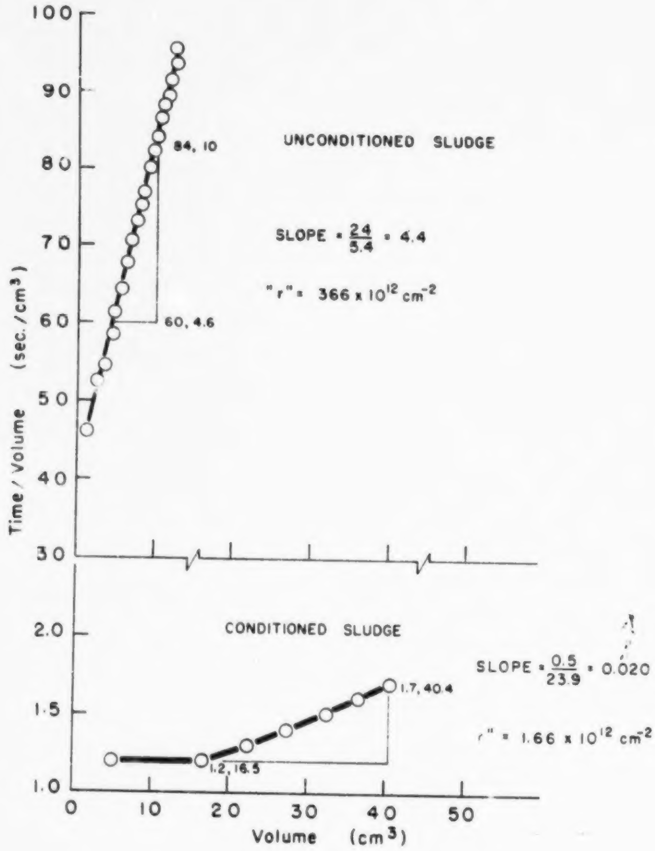

100
90
80
70
60
50
40
30
Time / Volume (sec./cm³)
84, 10
60, 4.6
UNCONDITIONED SLUDGE
SLOPE = 24/5.4 = 4.4
"r" = 366 x 10¹² cm⁻²
2.0
1.5
1.0
CONDITIONED SLUDGE
SLOPE = 0.5/23.9 = 0.020
r" = 1.66 x 10¹² cm⁻²
1.7, 40.4
1.2, 16.5
0
10
20
30
40
50
Volume (cm³)

Figure 33

SECTION IX

RESULTS

Sludge Characteristics

The source of digested sludges for dewatering was eight heated digesters operated in parallel. Each digester has a capacity of approximately 600,000 gallons. Six digesters have some sort of gas mixing while two gas holders have none. Hydraulic retention time of the digesters is approximately 12 to 16 days and the organic loading is approximately 0.08 to 0.10 pounds volatile solids per 1000 cubic feet.

Sludge feed to the digesters consists of raw sludge from primary clarifiers and secondary sludge from plastic media roughing filters and rock final filters.

Digested sludge is pumped more or less continuously to sludge holding tanks at the dewatering site. These holding tanks tend to stabilize the input to the dewatering process and provide some thickening.

Table II shows the average characteristics of the underflow of the sludge holding tanks to the process for the months of February through August 1972. The table shows that the solids concentration of the digested sludge was reasonably stable except for the months of June and July where large amounts of waste lime sludge from the water purification plant were received. During this period, the solids content increased to 7 percent and the organic content expressed as percent volatility decreased from 45 to 53 percent to 33 percent volatile.

Other parameters of digestion remained relatively constant. An increase in volatile acids was experienced in April due to a temperature recovery in the digesters after an extremely cold winter.

Filterability of Sludge

Digested sludge at Cedar Rapids is unusually difficult to dewater. As mentioned previously, European experience has indicated that a sludge with a specific resistance value of $150 \times 10^{12} \text{cm}^{-2}$ is an average sludge as far as the "degree of difficulty" in dewatering. Values of specific resistance for digested sludge at Cedar Rapids are all considerably higher than this value, typically possessing specific resistance values of 300 to $700 \times 10^{12} \text{cm}^{-2}$. Higher resistance values indicate greater difficulty in, or resistance to, dewatering.

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Table II

Average Characteristics Unconditioned Sludge
February to August 1972

	Feb.	Mar.	Apr.	May	June	July	Aug.	Av.
pH	7.1	7.2	7.2	7.3	7.3	7.4	7.5	7.28
Total Solids Percent	3.6	3.5	3.6	3.5	7.0	6.0	5.0	4.6
Volatile Solids Percent	53.8	51.9	53.2	45.0	33.6	36.9	44.7	45.6
Volatile Acids mg/l as Acetic Acid	49	41	410	62	50	77	73	108
Total Alkalinity mg/l as $CaCO_3$	4000	2900	2700	3600	3850	3800	4050	3557
Ammonia mg/l as N	846	594	604	838	944	993	1049	838
Organic Nitrogen as N	1464	1411	1338	1123	1272	1424	1493	1360
Grease- % of Total Solids	-	-	9.32	-	5.1	4.0	4.6	6.6

Complicating the picture is the fact that sludges at Cedar Rapids are not consistent. Digested sludge is always difficult but can be more or less difficult at various times.

One of the problems arises from the fact that the sludge varies considerably in the solids content. Because the value for specific resistance is inversely proportional to solids content, sludges with low solids content are generally more difficult to dewater than the same sludge which has been thickened. For example, if a 2 percent sludge, which has an r= 600 x $10^{12}cm^{-2}$ were thickened to 6 percent it would then have an r value of less than 200 x $10^{12}cm^{-2}$ providing, of course, that the basic nature of the sludge was not modified.

The basicpattern could be observed at Cedar Rapids as indicated in Table III.

Table III
Specific Resistance at $10^{12}cm^{-2}$

Percent Solids	r (Specific Resistance) x $10^{12}cm^{-2}$
2.43	772
3.33	571
5.34	342
6.68	236

As can be noted by the above table, the specific resistance is inversely proportional to the solids content. Unfortunately, however, in practice a 2 percent sludge cannot be handled as a mere volumetric dilution of a 4 to 6 percent sludge. Generally speaking, low solids sludge (2-3 percent) appears to contain a higher proportion of fines and requires a greater proportional amount of conditioning chemicals than a higher solids content sludge of 5.5 percent or greater.

Also it has been found that specific resistance of sludge may vary considerably at Cedar Rapids even though the percent solids remain relatively constant as shown in the following continuation of Table III.

Percent Solids	r (Specific Resistance) x $10^{12}cm^{-2}$
5.8	366
5.3	342
5.5	308
5.7	333
5.6	443
6.0	361

In the above table it can be seen that for the samples shown, while the percent solids are relatively uniform the specific resistance varied appreciably. For example, a 5.5 percent sludge had an $r = 308 \times 10^{12} cm^{-2}$ while a 5.6 percent sludge had an $r = 443 \times 10^{12} cm^{-2}$, a variance of 44 percent.

This variance in filterability is not unexpected considering the widely varying nature of the wastes as far as quality and quantity received at Cedar Rapids.

The most important ramifications of this varying nature of the sludge is that conditioning must

1. Be dependent on the solids content of the sludge, i.e., quantities (ash and chemicals) must vary as the percent solids vary.

2. Conditioning must be adjusted to the worst possible condition.

Because the conditioning charts are based on conditioning the worst possible sludge, a large quantity of sludge is often over-conditioned. In the data that will be presented later in this report, there will be instances where two sludges with the same solids content and essentially identical conditioning will have widely varied yields. The problem can be traced to the varying filterability as shown by specific resistance value in the table above.

Sludge Ash Characteristics

The source of sludge ash for dewatering was the ash from the multiple hearth incinerator used for the incineration of the cake from the pressure filter. Initial startup of the pressure filters was made with fly ash from a local power plant, consequently the initial characteristics of the sludge ash probably were a composite of fly ash and sludge ash, however, with the continued wasting of ash the sludge ash characteristics stabilized within a short period.

The physical and chemical properties of sludge ash are in all probability quite unique from installation to installation, depending upon the non-volatile portions of the sludge dewatered, chemical additives (lime, ferric chloride and/or polymers), temperature of combustion and the method of combustion (multiple hearth or fluidized bed).

At Cedar Rapids the sludge has at times a high inorganic fraction representing the discharge of waste lime from the lime soda softening process at the municipal water purification plant. No alum is used or discharged from the water purification plant.

Chemical additives in addition to sludge ash used for conditioning sludge are ferric chloride and lime.

The temperature of combustion ranges from 1200° F. to 1400° F.

Physical Properties of Sludge Ash

The physical properties of sludge ash are as follows:

Specific Gravity	2.63 - 2.78
Bulk Density	50 lbs. per cu. ft.
Color	Yellow
Mean Particle Size 29 mm	(22 - 40)

The particle size was determined by Process Measurements Section, Research Laboratory Branch, Control System Division of the Environmental Protection Agency with the cooperation of the Taft Center, Cincinnati. Particle size analyses were made using two different techniques, one being the electronic sensing of particle volume with the Coulter Counter Model T, and the second, aerodynamic sizing with the Bahco Microparticle Classifier.

In the following table are listed the average particle sizes of sludge ash as determined by the Coulter Counter Model T for nine samples of sludge taken from January 3, 1972 to May 10, 1972.

% of Particles Less than Given Size	Size in Millimeters x 10^{-3} Average of Nine Samples	Range mm x 10^{-3}
25	14.5	9.5 - 18
50	29.1	22 - 40
75	50.7	38 - 72
90	74.3	60 - 100

In Figure 34 is shown the results of a typical particle size determination by the Coulter Counter Model T.

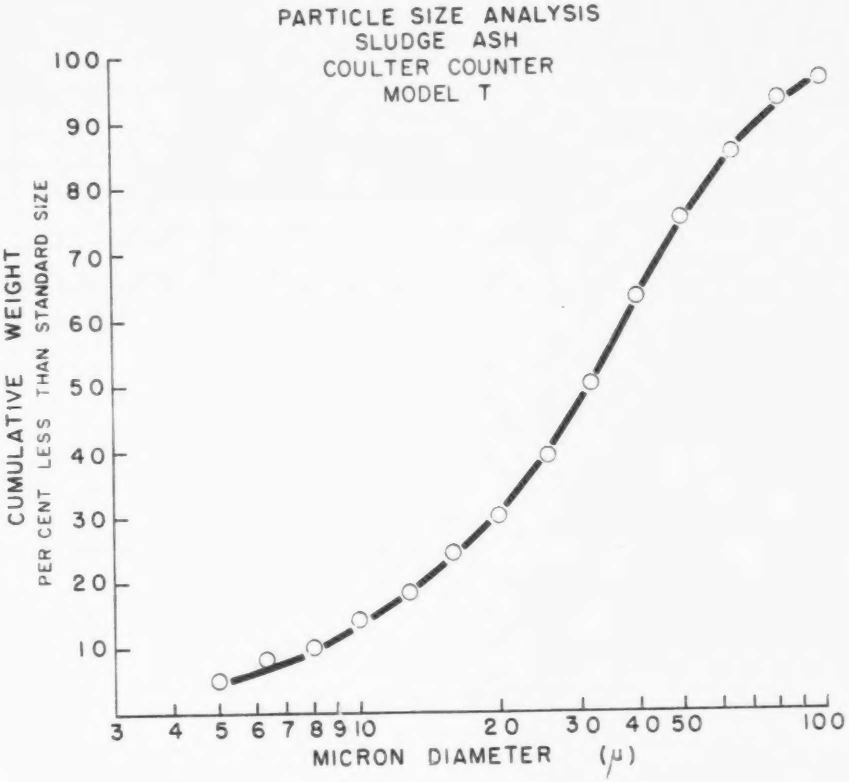


Figure 34

Particle size analysis performed by the Bahco Microparticle Size Analyzer are given in Table IV. Along with ash from the Cedar Rapids installation the table includes data on ashes from Mill Creek, Lake Tahoe, and Kansas City.

This table shows that for the incinerator ashes from Mill Creek, Lake Tahoe and Kansas City approximately 70 percent of the particles were 28 microns or less, for Cedar Rapids only 47.5 percent of the particles were less than 28 microns.

Table IV

Classification of Ash Particles
by Bahco Microparticle Size Analyzer

Effective Particle Size (10^{-3} mm)	Cedar Rapids %	Kansas City* %	Lake Tahoe* %	Mill Creek* %
0.00 - 0.92		1.49	1.38	2.22
0.92 - 1.60	3.0	2.95	2.71	3.70
1.60 - 2.97	5.0	5.70	7.53	7.81
2.97 - 7.71	12.5	9.05	9.83	21.86
7.71 - 12.19	9.5	14.56	10.95	26.52
12.19 - 20.75	10.5	22.41	16.20	22.46
20.75 - 24.72	4.0	9.21	10.13	6.83
24.72 - 27.30	3.0	8.83	4.28	8.60
28.38	52.5	25.8	36.99	

* Sludge Conditioning With Incinerator Ash
Smith, Hathaway, Farrell, Dean, AWTRL
EPA - Cincinnati, Ohio

Chemical Analysis of Sludge Ash

In Table V is shown the chemical analysis of sludge ash samples taken in January, May and September 1972. The samples are somewhat similar, however, the September ash sample had more calcium which would indicate the use of a greater amount of lime and less of the R_2O_3 group of oxides.

Table V

Chemical Analysis	May 24, 1972
Phosphate pentoxide, P_2O_5	4.41
Silica, SiO_2	22.96
Ferric oxide, Fe_2O_3	7.49
Alumina, Al_2O_3	6.27
Titania, TiO_2	0.31
Lime, CaO	50.60
Magnesia, MgO	1.62
Sulfur trioxide, SO_3	4.42
Potassium oxide, K_2O	0.78
Sodium oxide, Na_2O	0.59
Undetermined	0.55
	100.00

Silica Value - 27.77
T250 Poises: *1950° F.
* Less Than
Loss on Ignition at 1000° C. 16.50% wt. dry

Metals by Atomic Absorption -%	January 1972	September 1972
Ca	12.10	20.30
Fe	3.63	3.09
Mg	0.70	0.07
Zn	0.61	0.57
Cy	0.10	0.09
Cr	0.04	0.04
Cd	<0.01	0.00
Na	0.27	Not determined
K	0.40	Not determined
pH of 1% Solution	9.5	10.4

Fly Ash Characteristics

Fly ash for the project was obtained from a local power plant. The fly ash was the residue from combustion of pulverized coal. This residue consisted of a combination of ash collected by cyclone type dust collectors and electrostatic precipitators. The ash was transported dry from the power plant to the solidshandling plant by truck and pneumatically conveyed into the ash storage bins.

The physical properties of the ash are:

Color	Black
Specific gravity	2.3
Bulk density	43.6 lbs. per cu. ft.

Analysis of the particle size of fly ash was performed in the same manner as sludge ash using both the Coulter Counter and the Bahco type analysis. Coulter analysis of the particle size is shown below, and in Figure 35.

Percent Less Than Given Size	10^{-3} mm
25	11
50	20
75	33
90	42

PARTICLE SIZE ANALYSIS
BAHCO ANALYZER

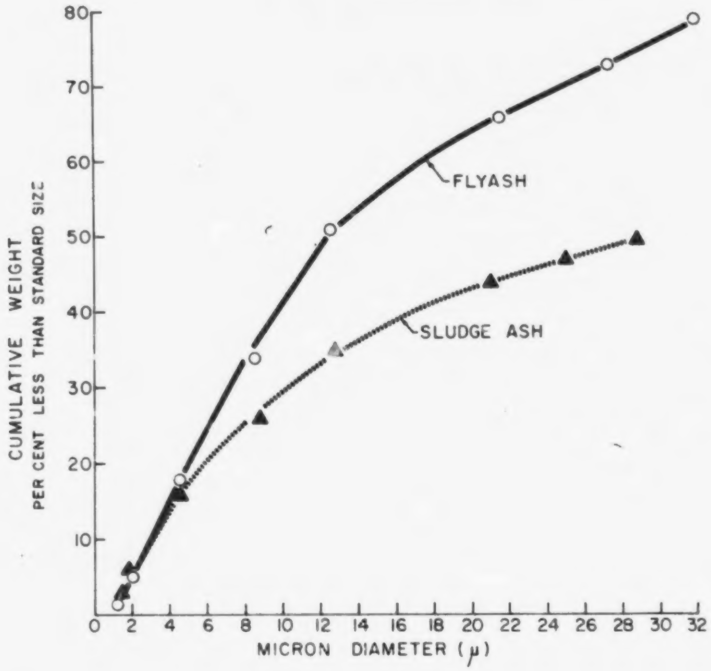


Figure 35

Fly ash particle size was also determined by using the Bahco apparatus. Results are shown in Table VI.

Table VI

Classification of Fly Ash Particles
by Bahco Microparticle Size Analyzer

Effective Particle Size (10^{-3} mm)	Percent
0.00 - 1.60	2.5
1.60 - 2.97	7.5
2.97 - 7.71	20.5
7.71 - 12.19	16.5
12.19 - 20.75	17.5
20.75 - 24.72	5.5
24.72 - 28.30	5.0
28.30	25

Chemical analysis of fly ash is shown in the following table.

TABLE VII

Chemical Analysis of Fly Ash

Phosphate pentoxide, P_2O_5	0.85
Silica, SiO_2	53.90
Ferric oxide, Fe_2O_3	17.00
Alumina, Al_2O_3	11.93
Titania, TiO_2	0.57
Lime, CaO	6.80
Magnesia, MgO	1.65
Sulfur trioxide, SO_3	2.30
Potassium oxide, K_2O	4.32
Sodium oxide, Na_2O	0.39
Undetermined	0.29
	100.00

Silica Value - 67.93
T250 Poises: 2355° F.
Loss on Ignition at 1000°C. 14.5%

As can be noted from Table VII, fly ash is mainly composed of silica, iron oxide and alumina.

Comparison of Sludge Ash and Fly Ash as a Conditioning Agent for Sludge

Extensive testing was performed using the resistance meter in comparing the effects of sludge ash and fly ash as a conditioning aid for the pressure filtration of sludge, both digested sludge and raw sludge.

The reason for the initial testing was to attempt to explain the apparent lack of correlation between the pilot plant study and the performance of the full scale filter. The pilot plant had operated without the use of any other conditioning agent except fly ash, the full scale plant could not successfully operate at any reasonable sludge/ash ratio without using ferric chloride and lime.

Consequently a program of study was set up comparing the two different ashes. Experimentally the ashes were compared by securing a sample of digested sludge, determining the solids content of the sludge and then dosing liter samples with varying amounts of ash, up to an ash ratio of 3 to 4.

In Figures 36 to 40 are shown an example of the testing. Figure 36 represents a sample of digested sludge with a solids content of 2.5 percent. The initial specific resistance of the sludge was 772. At an ash/sludge ratio of 1.3 with fly ash the specific resistance was lowered to 78 x 10^{12} cm^{-2} while at an ash/sludge ratio of 3.95 for sludge ash the specific resistance was 67 x $10^{12}cm^{-2}$ comparable values.

Figure 37 represents a sample of digested sludge with a solids content of 5.5 percent and an initial specific resistance of 308 x $10^{12}cm^{-2}$. In this case an ash/sludge ratio of 1.5 for fly ash gave a specific resistance of 12.3 x $10^{12}cm^{-2}$ while an ash/sludge ratio of 4.0 for sludge ash gave a specific resistance of 14.1 x $10^{12}cm^{-2}$, again essentially the same values.

Figure 38 represents a sample of digested sludge with a solids content of 5.6 percent and a specific resistance of 443 x $10^{12}cm^{-2}$. Here a value of 1.0 ash/sludge ratio for fly ash has a specific resistance of 63.2 x $10^{12}cm^{-2}$, comparable to a value of 61.0 x $10^{12}cm^{-2}$ for an ash/sludge ratio of 2.5 for sludge ash.

The above data indicates quite strongly that fly ash is 2.5 to 3.0 times more effective for conditioning digested sludge per unit weight of ash than sludge ash.

Additional work was performed in comparing the conditioning effect of the two ashes with other sludges. If the ashes performed in the same manner with other sludges then any individual peculiarities of the digested sludge at Cedar Rapids would be ruled out.

Figure 39 shows a comparison of fly ash and sludge ash as conditioning agents for raw sludge at Cedar Rapids. Here again it can be seen that fly ash was consistently better than sludge ash.

Figure 40 shows a comparison of fly ash and sludge ash as conditioning agents for a digested sludge obtained from a waste water treatment plant operated by the City of Cedar Rapids which handles strictly domestic sewage. Here again fly ash performs better than sludge ash.

In summary, the fly ash obtained from a local power plant at Cedar Rapids performs significantly better as a sludg conditioning agent than ash obtained from the incineration of sludge at Cedar Rapids.

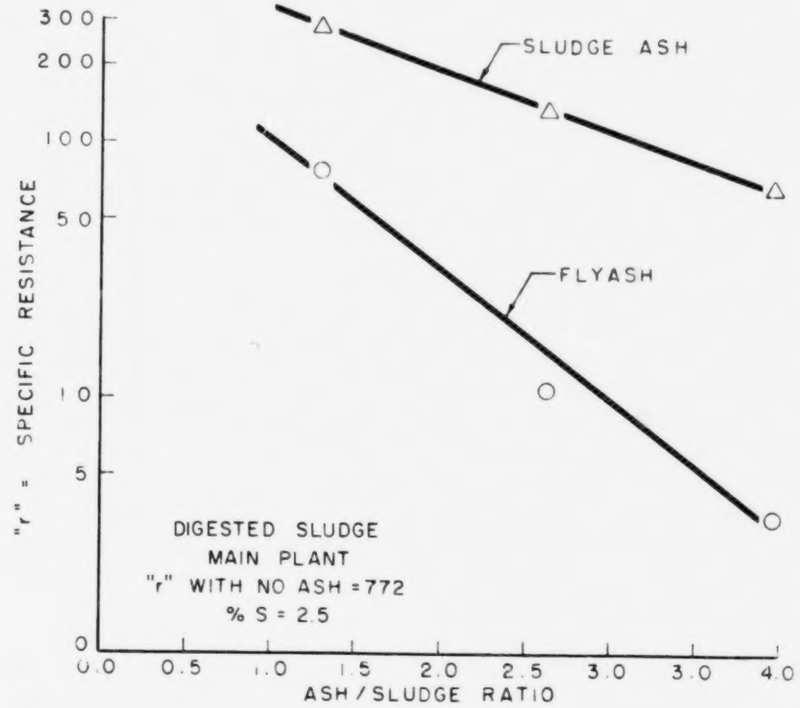


Figure 36

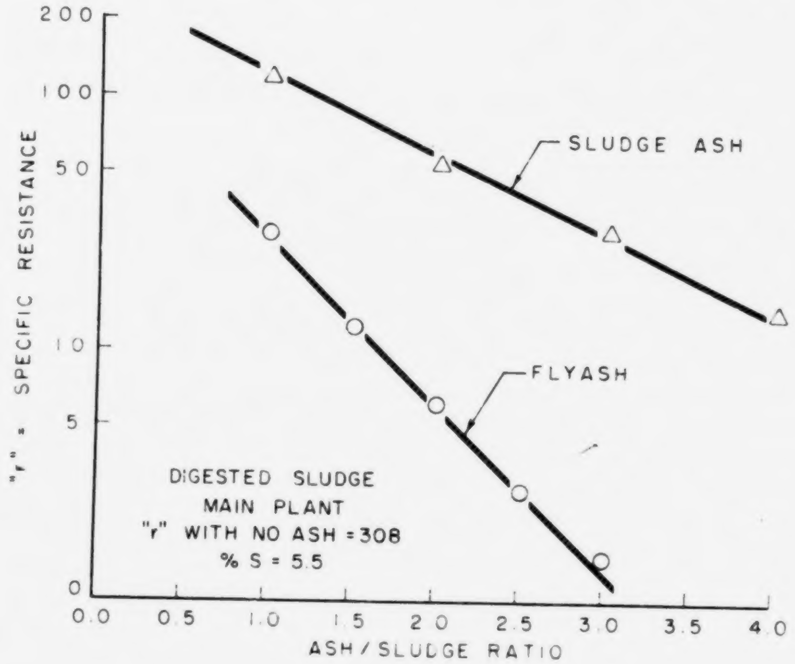


Figure 37

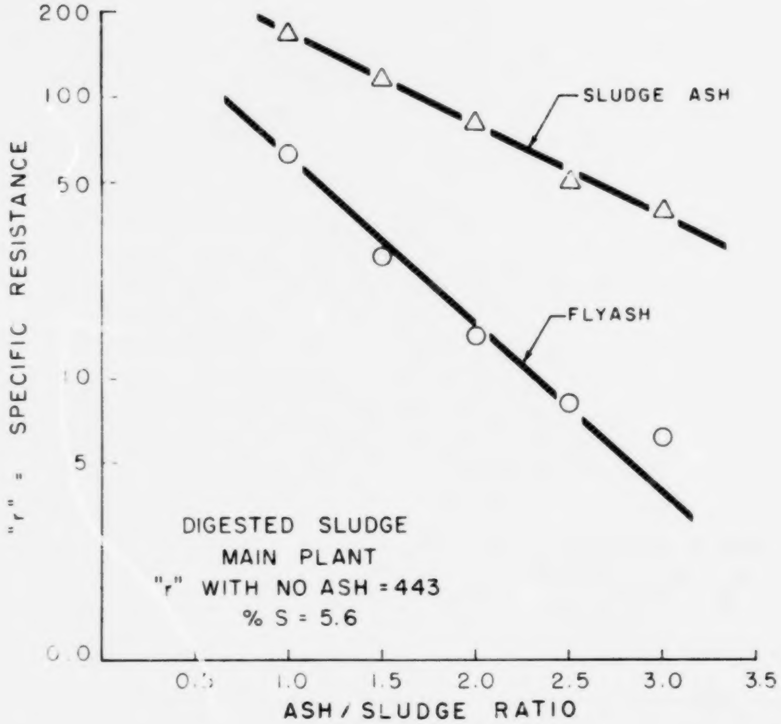


Figure 38

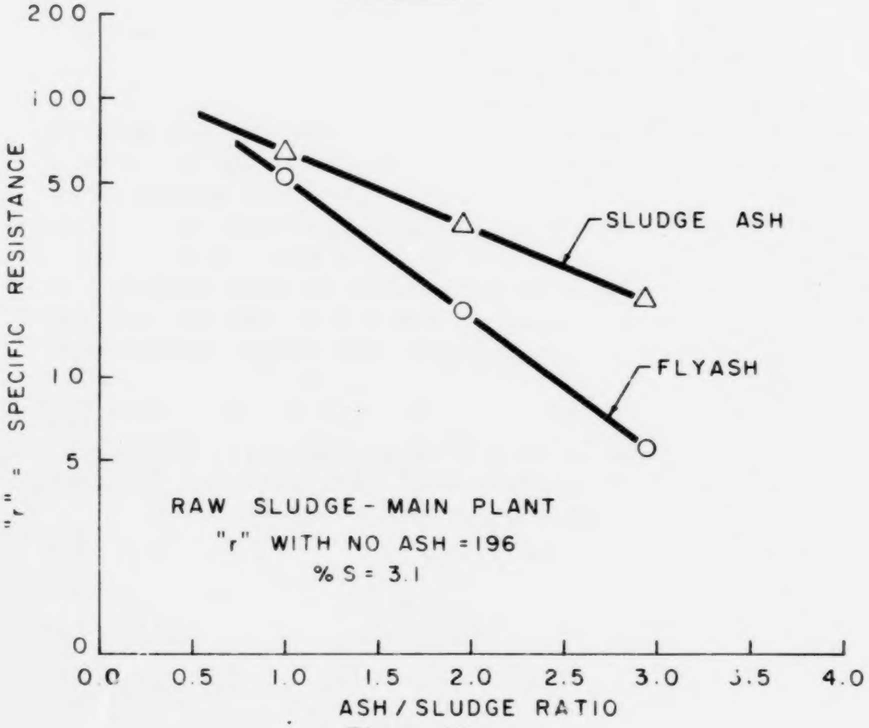


Figure 39

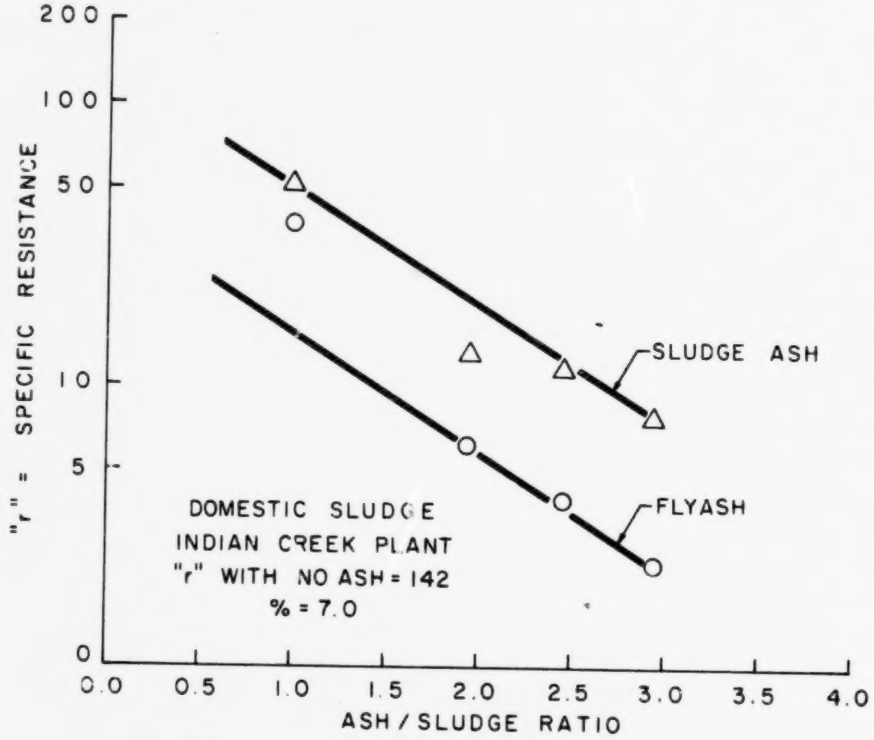


Figure 40

As previously stated, a substantial difference exists between the performance of fly ash and sludge ash as a conditioning agent for sludge. Significantly this is a problem which is unique at Cedar Rapids. Studies by others have found differences between sludge ash and fly ash but in reverse order, i.e., fly ash having little effect or sludge ash being much better. At Cedar Rapids the fly ash is a good conditioning agent significantly better than sludge ash, therefore, it appears that the fly ash and possibly the sludge ash at Cedar Rapids are unique and results cannot be directly applied to other cities.

Fly ash and sludge ash can differ in three fundamental ways, chemically, by size, and by shape. Chemical analyses have been presented for both sludge ash and fly ash but are given in Table VIII for comparison.

Table VIII

Comparison of Chemical Analysis of Fly Ash and Sludge Ash

Results % by Weight on Ignition at 1000° C.

		Fly Ash	Sludge Ash
Phosphorous pentoxide	P_2O_5	0.85	4.41
Silica	SiO_2	53.90	22.96
Ferric oxide	Fe_2O_3	17.00	7.49
Alumina	Al_2O_3	11.93	6.27
Titania	Ti_2O_3	0.57	0.31
Lime	CaO	6.80	50.60
Magnesia	MgO	1.65	1.62
Sulfur trioxide	SO_3	2.30	4.42
Potassium oxide	K_2O	4.32	0.78
Sodium oxide	Na_2O	0.39	0.59
Undetermined		0.29	0.55
		100.00	100.00

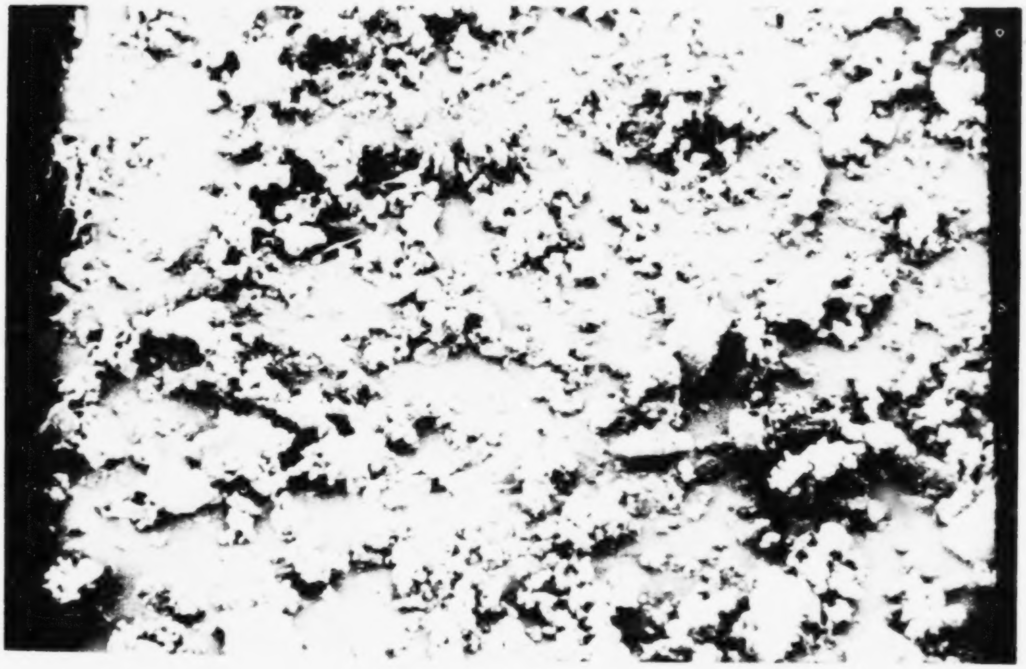
Chemically fly ash and sludge ash are quite dissimilar. Fly ash is approximately 50 percent silica and sludge ash is approximately 50 percent calcium oxide. Other components such as ferric oxide and alumina are present in differing quantities. The effect of these higher levels of iron and aluminum salts in fly ash is not known and deserves further research.

Particle Size

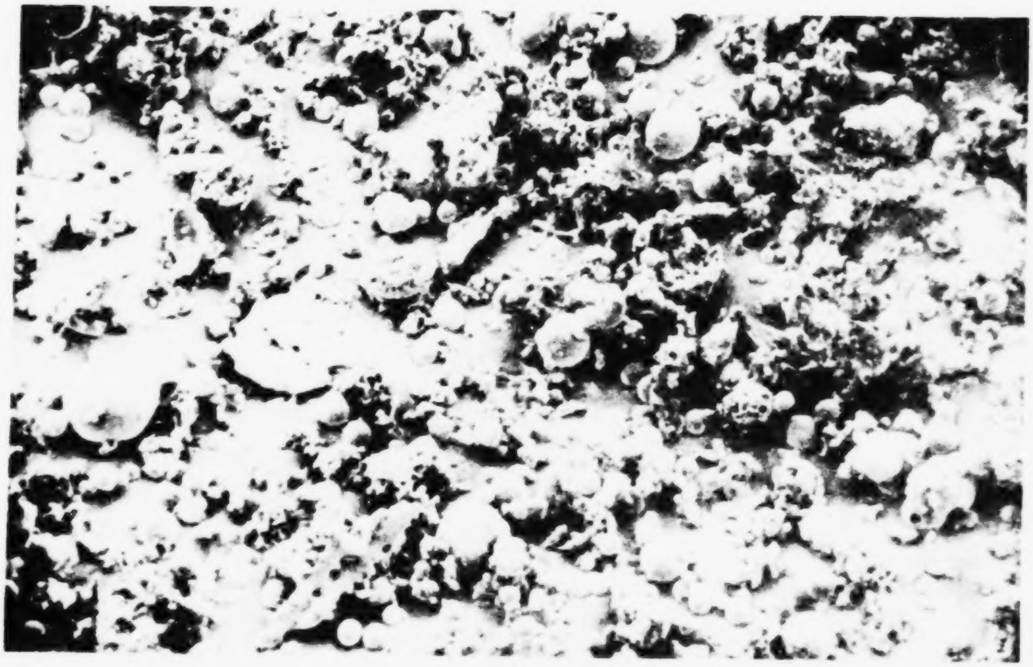
The range of particle sizes for fly ash and sludge ash have been given in previous sections. Sludge ash particle analysis is shown in Figure 34, and fly ash analysis is shown in Figure 35. Fly ash has a smaller particle size than does sludge ash. As has been shown in previous sections, ash fractions of smaller size ($< 45\mu$) are significantly far more effective for sludge conditioning than ash of larger size($> 45\mu$). Further research is needed to further clarify the effect of particle size on sludge filtration.

Particle Shape

Photo micrographs of sludge ash and fly ash were obtained by a scanning electron microscope.

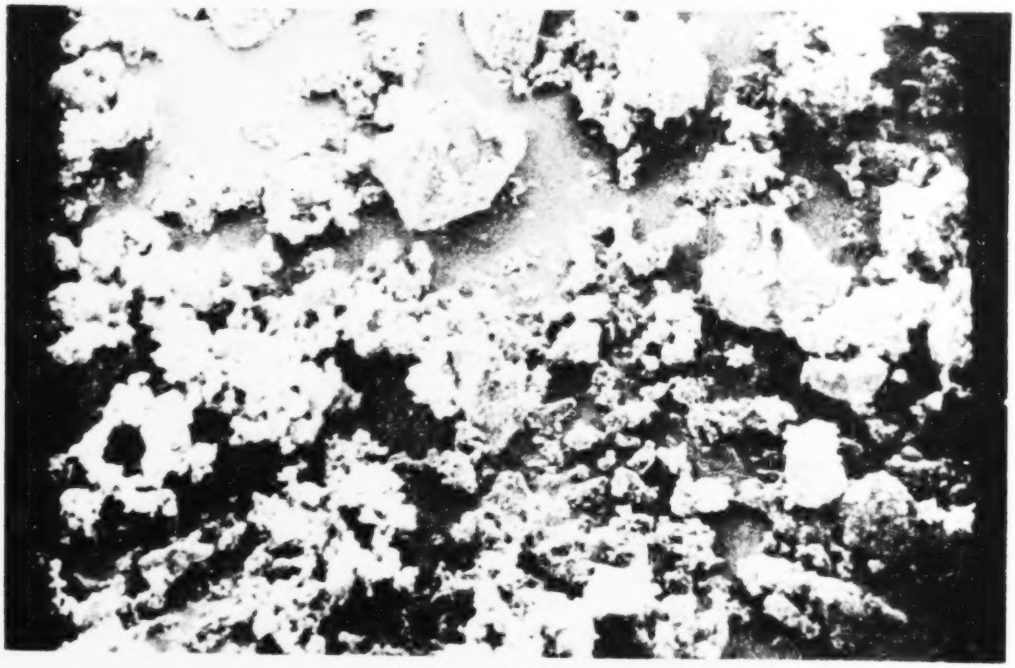


Sludge Ash 200X

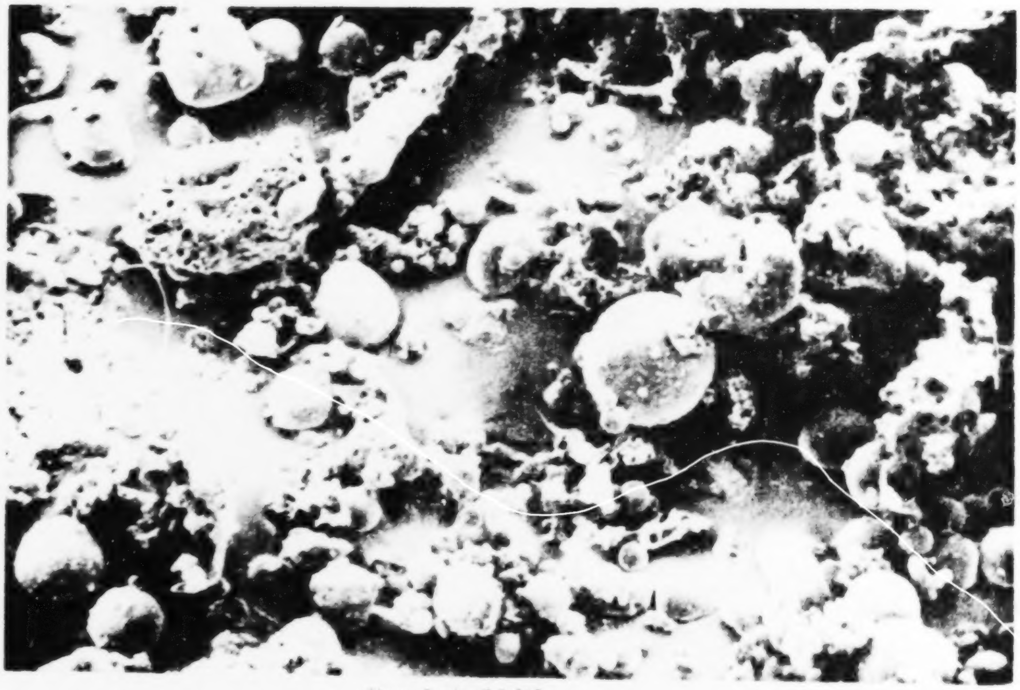


Fly Ash 200X

Figure 41

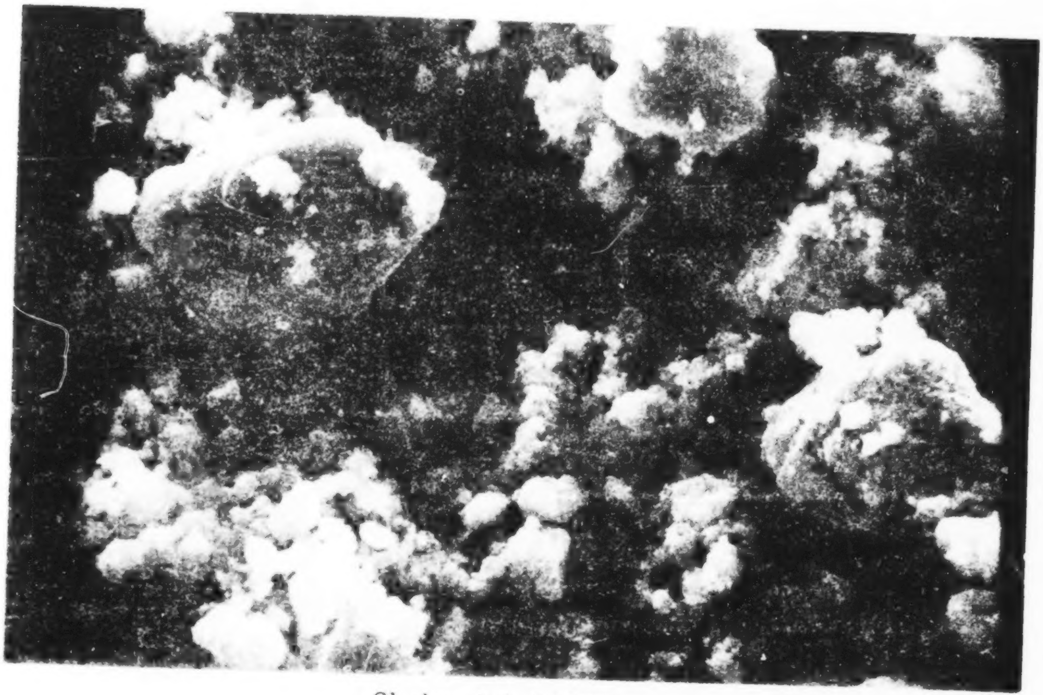


Sludge Ash 500X

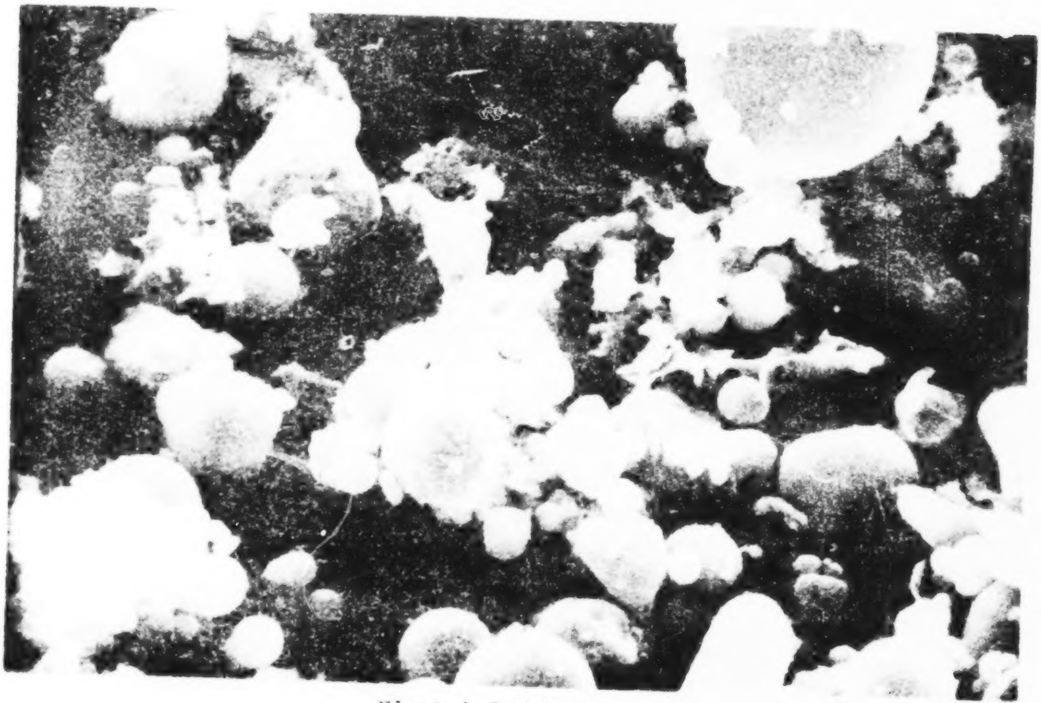


Fly Ash 500X

Figure 42



Sludge Ash 1000X



Fly Ash 1000X

Figure 43

A small amount of each sample was vacuum coated with a thin conductive layer of first carbon, then gold. The coated samples were then examined by a scanning electron microscope. On the preceding pages are comparison photographs of fly ash and sludge ash taken at magnifications of 200x, 500x, and 1000 x. Figures 41, 42 and 43.

From these photomicrographs it appears that fly ash tends to be more spherical than sludge ash. The sludge ash is more irregular and appears to agglomerate. In comparing particles of each ash at equal magnifications, fly ash appears larger than sludge ash although both Coulter Counter and Bahco Microparticle Classifier show otherwise. This apparent particle size problem may be attributed to the fact that the sludge ash agglomerates. In laboratory work the sludge ash tends to clump in water and resists dispersion much more than does fly ash. This tendency would be characteristic of a material which agglomerates.

Effect of Particle Size on Specific Filter Resistance

Concern has been expressed that, in sludge filtration processes using recycled sludge incinerator ash as a filter aid, that the recycling of ash through the incinerator may result in a shift of ash particles to the smaller particle sizes. This concern is based upon the premise that the fine fraction of ash is not effective in sludge conditioning and may in fact be deleterious.

Consequently sludge ash samples from Cedar Rapids were split up on an air jet sieve into the following fractions.

0 - 15 microns
15 - 32 microns
32 - 63 microns
63 - 90 microns

These fractions were subsequently used to condition a digested sludge from Wiesbaden Biedrich which had a specific resistance of $159 \times 10^{12} \text{cm}^{-2}$. All the samples were mixed with ash equivalent to a 1.0 ash/sludge ratio, or 42.1 grams ash per liter, for this 4.21 percent solids sludge. These ash/sludge mixtures were then tested for specific filter resistance. The following results were obtained.

Fraction (microns)	A/S	$r \times 10^{12} \text{cm}^{-2}$
0 - 15	1.0	29.4
15 - 32	1.0	46.5
32 - 63	1.0	62.4
63 - 90	1.0	67.0
Original mixture all fractions	1.0	60.4

Here it was observed that a decreasing ash particle size enhances the ability of the ash to condition the sludge. Sample conditioned with fractions of ash below 32 microns had less filter resistance than the original mixture and the most effective fraction of ash was that fraction below 15 microns.

The preceding data was obtained in Germany by Dr. O. Busse, Passavant Werk, during an earlier evaluation of the Cedar Rapids process.

In Cedar Rapids the sludge incinerator ash was mechanically separated into three fractions.

0 - 45 microns
45 - 106 microns
> 106 microns

These fractions of sludge ash were used to condition digested sludge from Cedar Rapids. Ash to sludge ratios of 2.1 were prepared and then tested for specific resistance.

Type of Ash	Size	A/S	r
None		0	467
Sludge ash	0- 45 microns	2.1	77.1
Sludge ash	45-106 microns	2.1	110
Sludge ash	> 106 microns	2.1	112
Sludge ash	All sizes (Original)	2.1	88.3
Fly ash	All sizes	2.1	44.7

These results at Cedar Rapids verify the work done in Germany. Again the smaller fractions of ash are more effective in conditioning sludge. However, the smallest fraction of sludge ash is still not as effective as fly ash as indicated above. Unfortunately, equipment was not available for fractionating sludge ash below 45 microns. It may be possible that smaller fractions of sludge ash may have performed as well as fly ash since 78 percent of fly ash has a particle size less than 32 microns.

It appears that diminution of ash with the subsequent increases in the fraction of fines with recycle through an incinerator is not a problem. The opposite condition of the fines being eliminated might pose a real problem. Operation with recycle of incinerator ash at Cedar Rapids has not apparently had a significant effect upon ash particle sizes.

Conditioning Digested Sludge with Sludge Ash and Chemicals

Quite early in the operation of the pressure filter it became apparent that the digested sludge could not be conditioned satisfactorily with sludge ash alone. Consequently, in January of 1972 Dr. Oswald Busse of the Passavant Werk, West Germany, came to Cedar Rapids to assist in a series of tests to determine how the digested sludge could be conditioned for satisfactory filter performance.

Dr. Busse's testing program consisted of dosing various samples of sludge with ash, ferric chloride, and lime and determining the specific resistance of the conditioned sample. Adequate conditioning was reached when the conditioned sludge sample had a specific resistance below $r = 20 \times 10^{12} cm^{-2}$.

Based upon this testing program and previous experience, Dr. Busse made the following recommendations for ash, ferric chloride and lime dosages which are still basically in use today.

Figure 44 shows the dosage chart used for operation. Dosages are based upon pounds of conditioning agent (lime, ferric chloride or ash) per 100 gallons. The dosages are presented in this manner because the sludge is generally pumped from the holding tanks through conditioning to the contact tank at 100 gallons per minute.

Ferric chloride dosages range from 2.2 pounds liquid (0.88 pounds actual) per 100 gallons at 2 percent solids to 3.3 pounds liquid (1.32 pounds actual) per 100 gallons at 15 percent solids. These dosages on a pound of dry ferric chloride per ton dry solids basis range from 103 pounds per ton at 2 percent solids to 21 pounds per ton at 15 percent solids. Ferric chloride costs at Cedar Rapids are $0.0625 per pound delivered.

Lime dosages range from 6.2 pounds per 100 gallons at 2 percent solids to 7.5 pounds per 100 gallons at 15 percent solids. This is equivalent to 743 pounds per ton dry solids at 2 percent to 120 pounds per ton dry solids at 15 percent. Lime costs at Cedar Rapids are $0.0114 per pound.

Another way of expressing the recommended ash quantities for properly conditioned sludge is shown in Figure 45. It is apparent that below a sludge concentration of about 5.5 percent the ash/sludge ratio must be increased in relation to the decreased sludge concentration to assure a properly conditioned sludge for filtration.

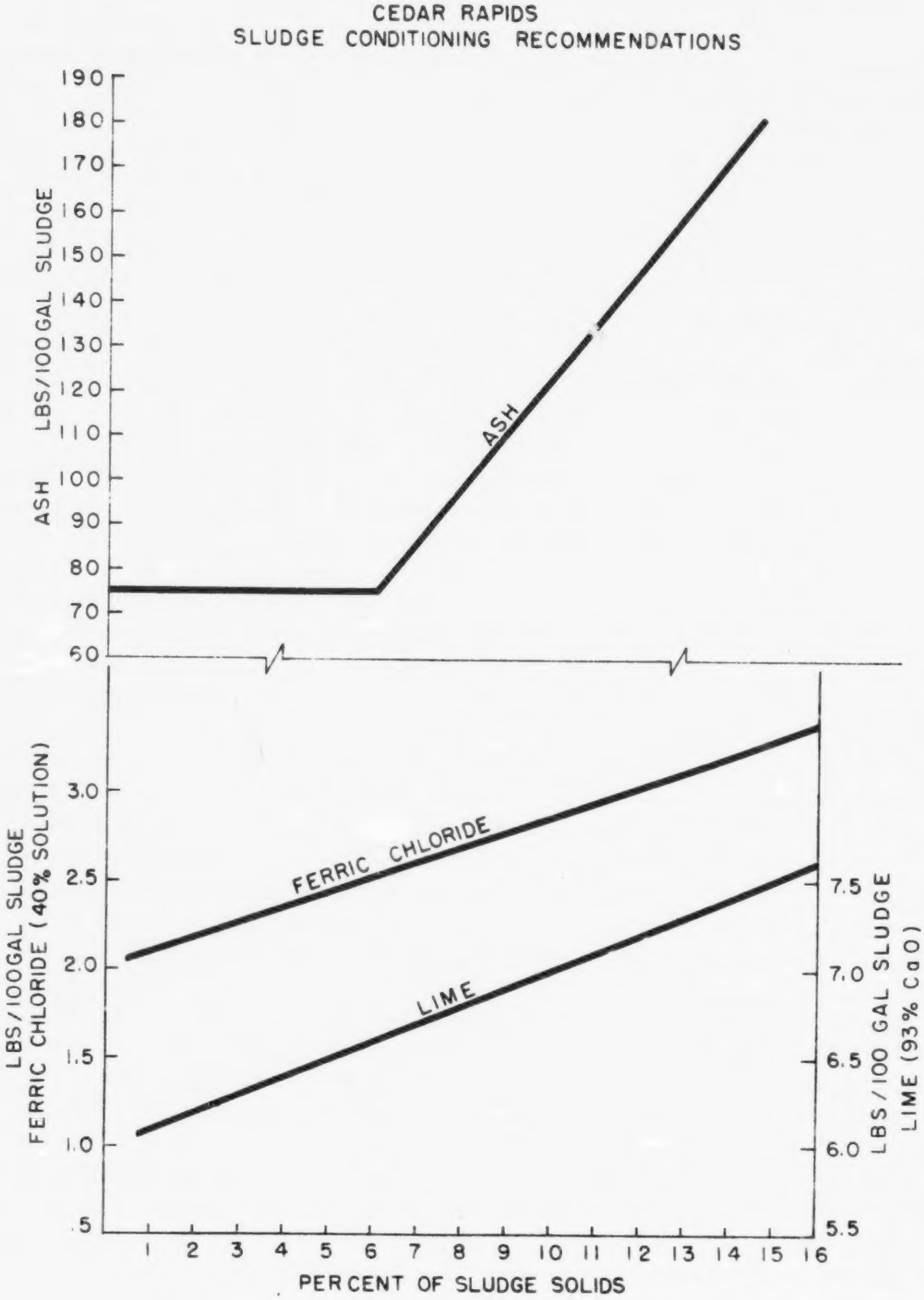

CEDAR RAPIDS
SLUDGE CONDITIONING RECOMMENDATIONS
ASH LBS/100 GAL SLUDGE
190
180
170
160
150
140
130
120
110
100
90
80
70
60
ASH
LBS/100GAL SLUDGE FERRIC CHLORIDE (40% SOLUTION)
3.0
2.5
2.0
1.5
1.0
.5
FERRIC CHLORIDE
LIME
LBS/100 GAL SLUDGE LIME (93% CaO)
7.5
7.0
6.5
6.0
5.5
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
PERCENT OF SLUDGE SOLIDS

Figure 44

Equivalent costs of these chemical dosages are shown in Figure 46. Chemical costs range from $14.89 per ton at 2 percent solids to $2.69 per ton at 15 percent solids. The optimum solids concentration appears to be near 6 percent. At this point the cost per ton for chemicals is $5.48 and the ash/sludge ratio is at a minimum 1.5/1. These dosages and costs are essentially theoretical and will be compared to actual costs later in the report.

These costs also are not necessarily typical for sludge dewatering with pressure filtration but are dependent upon specific resistance of the sludge and the nature of the sludge ash.

Filter Performance Curves

Filter performance curves were established plotting the filtration rate (yield) in terms of pounds per hour per square foot, versus time in hours at various sludge solids concentrations. Figures 47 to 56 show filter performance for feed sludge solids ranging from 2.5 percent to 6.5 percent.

Sludge conditioning requirements were determined with bench studies using the resistivity meter prior to actual filtration in the pressure filter. It was obvious that proper sludge conditioning was necessary to assure good filter operation and a satisfactory cake product. On the other hand, if good sludge conditioning was obtained then the process performed very well at a given level of production and that production was fairly predictable over a broad range of raw solids input. Referring to the filter performance curves, Figures 47 to 55, it is apparent that production (yield) has an optimum ash/sludge ratio for each sludge density. The optimum yield is at the lowest ash/sludge ratio which will assure good sludge conditioning as determined on the bench using the resistivity meter. As an example, at 5.5 percent solids, an ash/sludge ratio of 1.1/1 is slightly better than an ash/sludge ratio of 1.5/1 as shown in Figure 53. However, as the sludge density is decreased to 2.5 percent solids, the ash/sludge ratio is necessarily increased to 4.1/1 as shown in Figure 47 and the optimum ash/sludge ratio is still the lowest ratio which will provide good sludge conditioning.

Maintaining good sludge conditioning, the yield, expressed as pounds per hour per square foot of filter media increases as the sludge solids concentration increases as indicated by Figure 56, which is a best-fit composite curve from all of the computer curves of Figures 47 to 55.

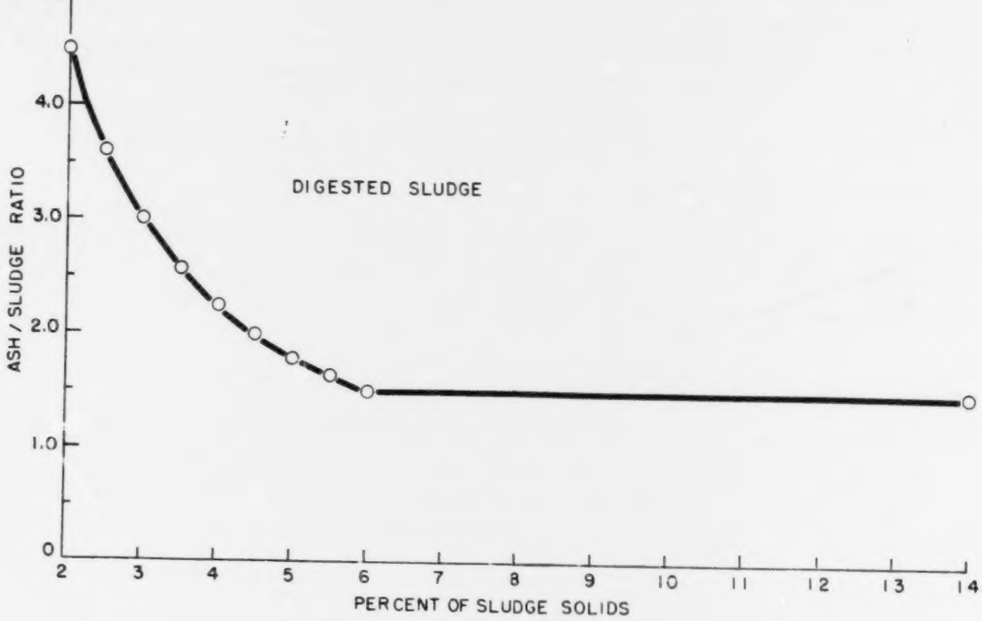


Figure 45

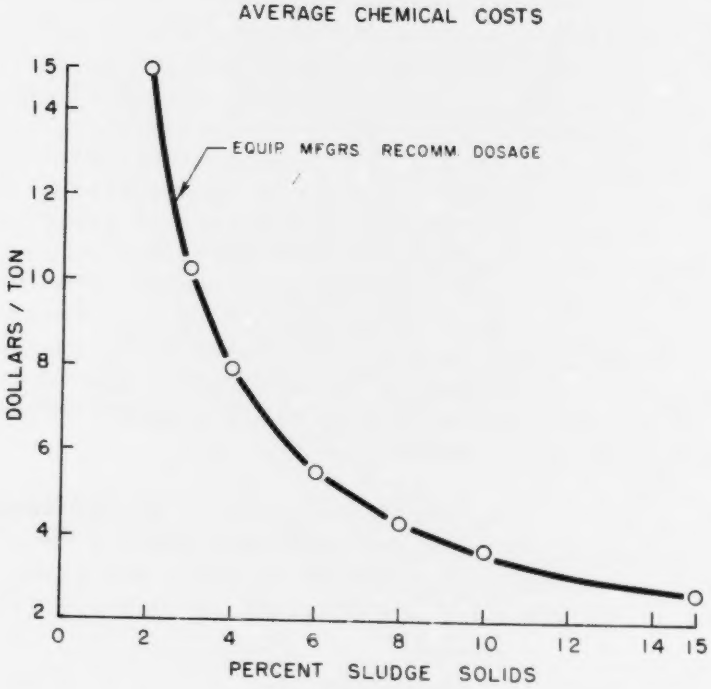
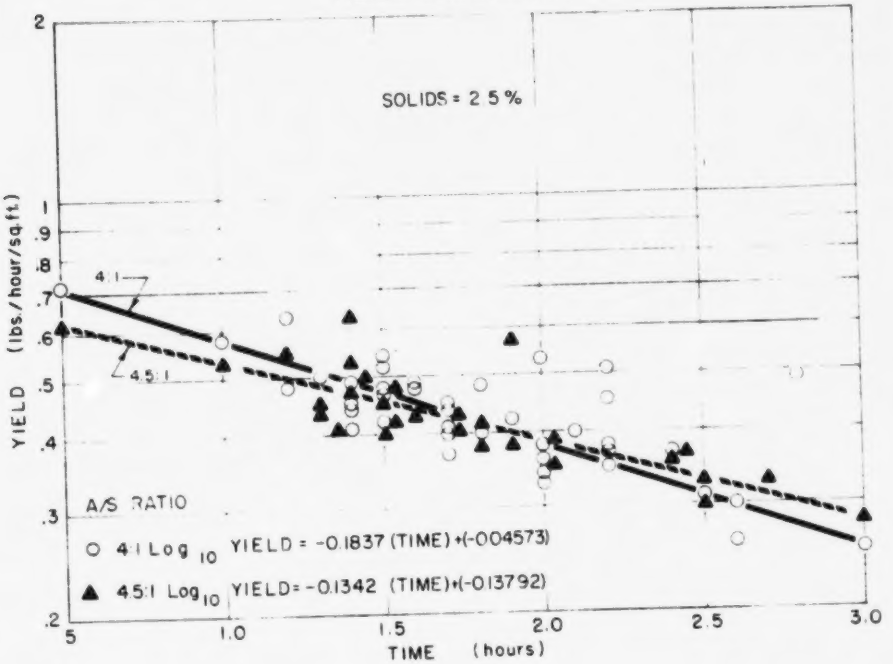
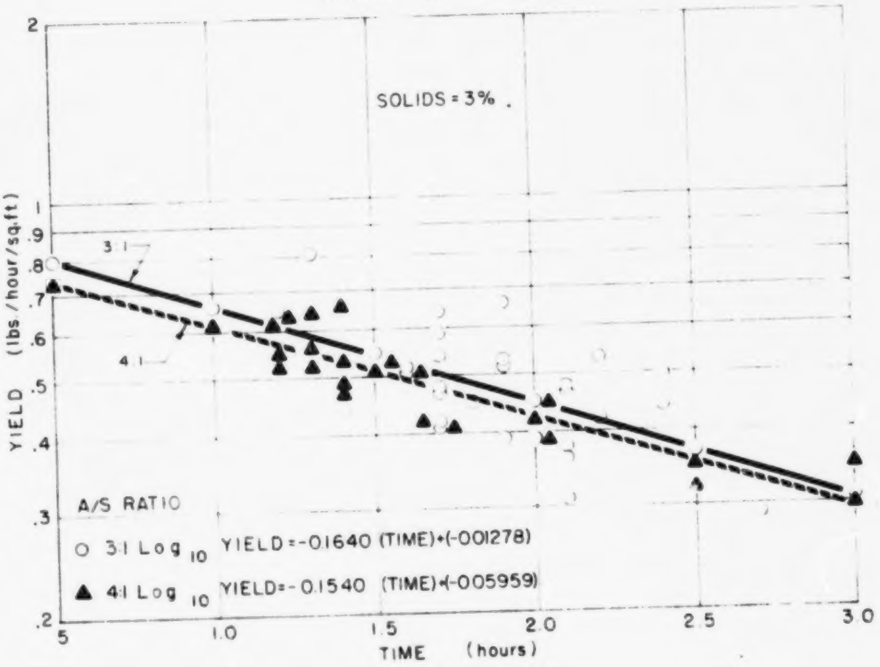


Figure 46



FILTER PERFORMANCE

Figure 47



FILTER PERFORMANCE

Figure 48

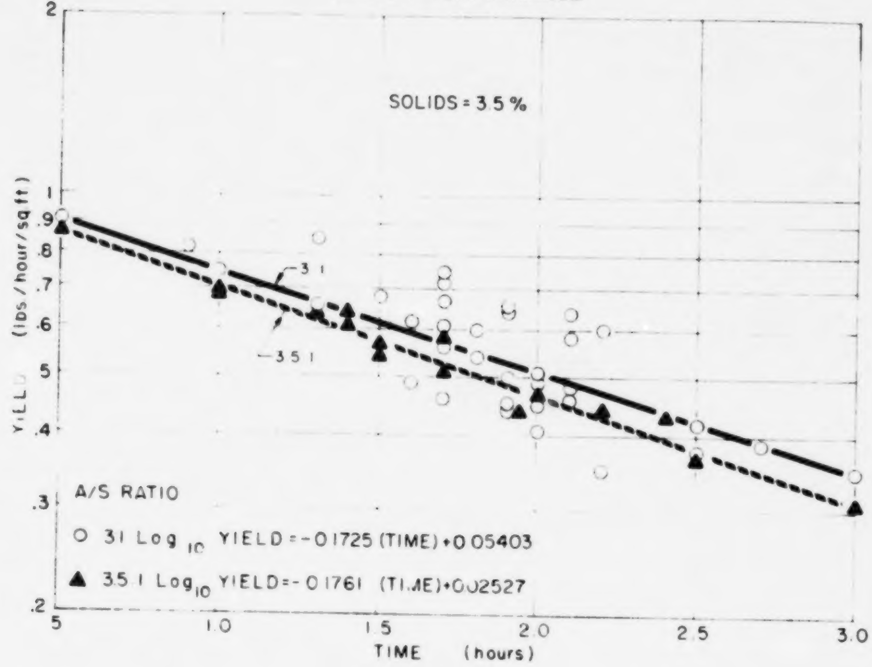


Figure 49

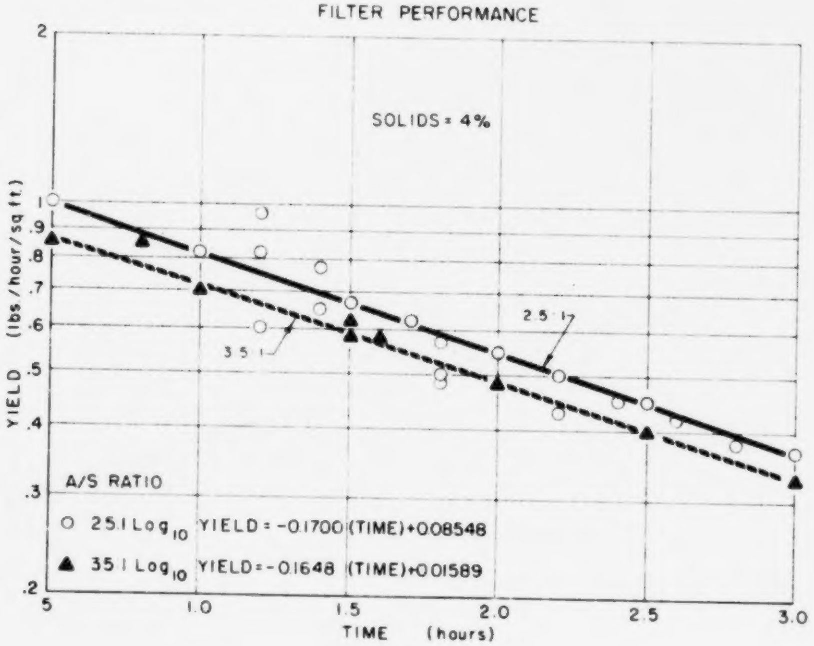


Figure 50

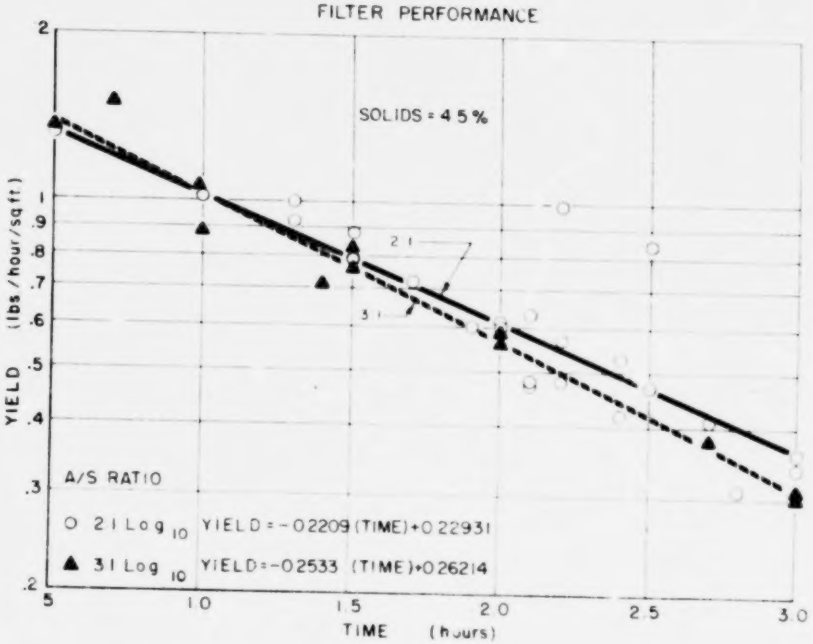


Figure 51

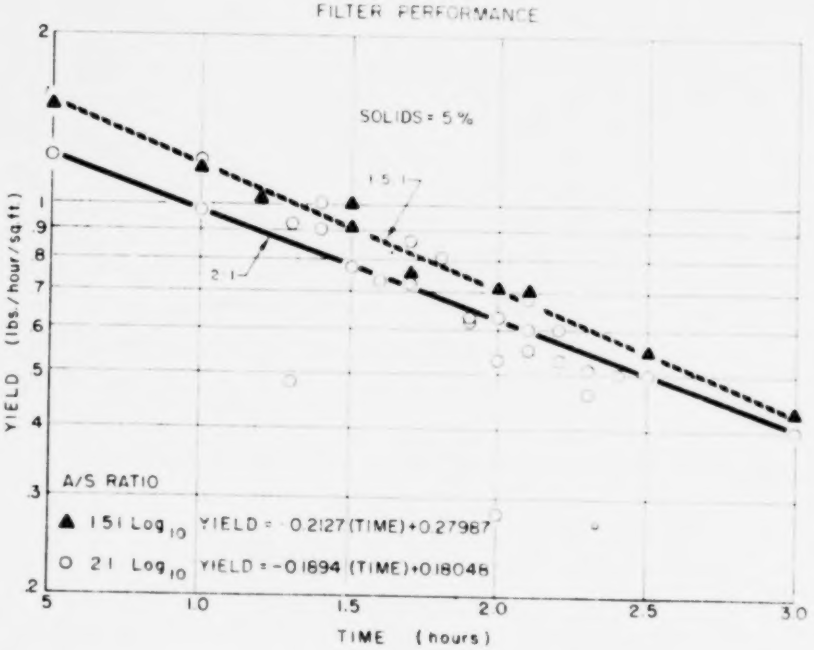


Figure 52

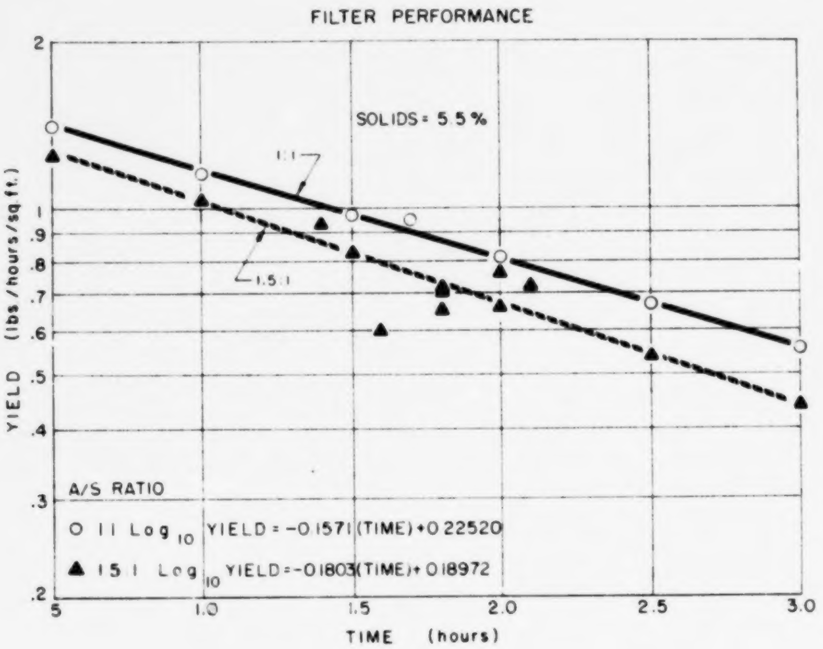


Figure 53

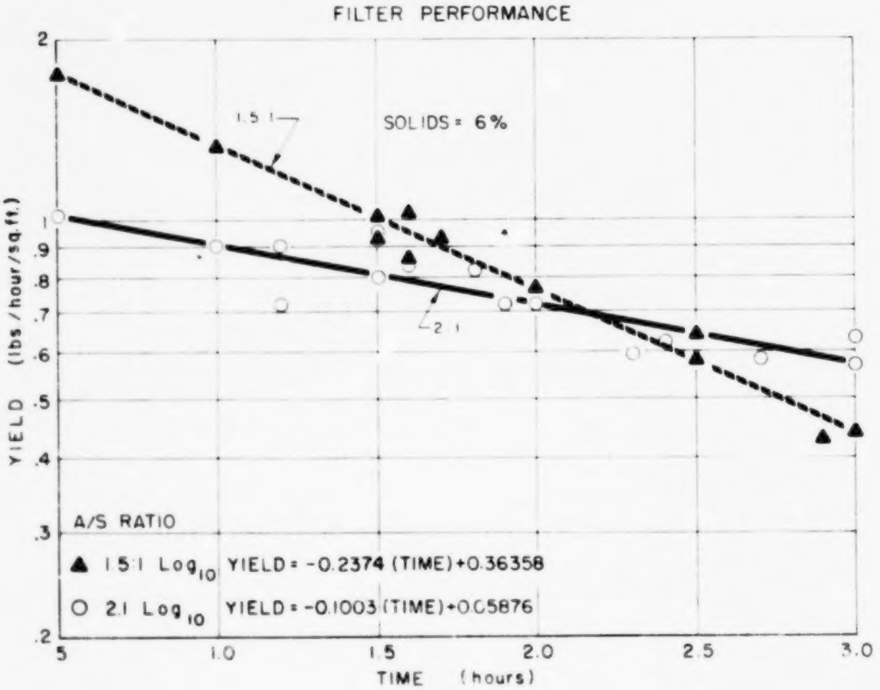


Figure 54

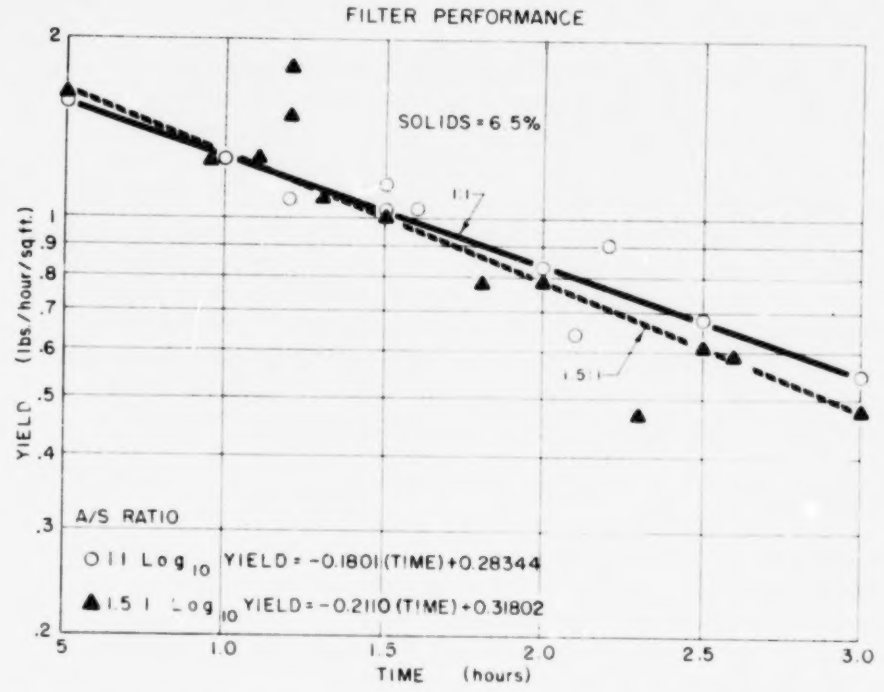


Figure 55

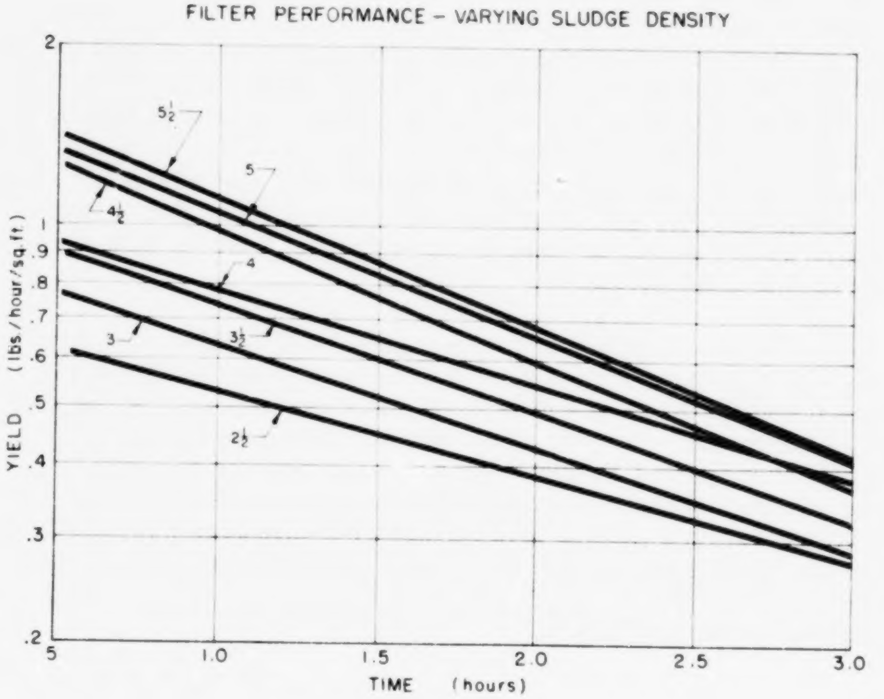


Figure 56

Filter yield increases in almost direct proportion to increased raw sludge solids density at a ratio of about 1.8/1, that is, if the sludge density is increased times 2, the resulting filter yield is increased times 1.8.

It is apparent that properly conditioned sludge filters satisfactorily at both low and high sludge densities and that yield is principally influenced by time to completely fill the cavity forming the predetermined cake volume. In practice it became apparent that the limiting factor to improving yield was the ability to pump conditioned solids to the pressure filter in sufficient capacity to develop the maximum rate of filtration. Each system consisted of two pumps and some experimentation was performed using three and four pumps simultaneously on one system.

Filtrate Characteristics

The pressure filter is a very efficient solids liquid separator with efficiencies of separation practically at 100 percent. During the months of February to August 1972 the filtrate averaged 74 mg/l suspended solids while the sludge solids averaged 4.60 percent solids giving a percent removal of suspended solids of 99.99984 percent.

During the pilot study it was noted that during a run the suspended solids in the filtrate were relatively high initially (400 mg/l approx.) and then decreased to very low values during the run. No variance in suspended solids could be noted during a run on the full scale filter. Possible explanations for this difference may lie in two areas, the initial pressure on the pilot plant was a full 225 psig while on the full scale filter the pressure started at a low value (approximately 40 psig) and gradually increased to 225 psig. The full 225 psig pressure may have forced suspended solids through the thin precoat on the pilot plant while this did not occur on the full scale filter.

The BOD and COD of the filtrate are due primarily to dissolved organics such as volatile acids and are not necessarily affected by the filtering process. It is possible that some dissolved COD may be removed by the process of absorption, however, this was not investigated on the full scale filter and deserves further research. Limited bench work indicates that conditioning with fly ash alone may remove as much as 50 percent of the dissolved COD. This removal was not noted in samples conditioned with sludge ash alone.

Filtrate quality is shown in Table IX.

Table IX

Average Filtrate Quality

	Feb.	Mar.	Apr.	May	June	July	Aug.	Average
Suspended Solids mg/l	25	94	103	60	84	65	88	74
Total Solids mg/l	5000	4900	5600	6200	6400	6000	6800	5842
Volatile Percent Solids	11.2	15.6	23.7	20.8	26.0	30.2	25.1	21.8
Turbidity	18	20	29	19	17	22	27	21.7
Color Cobalt Platinum Units	62	61	69	89	107	123	132	91.8
Conductivity Micromhos/cm^2	7883	7520	7914	7483	6570	5800	5600	6967
COD mg/l	623	706	1310	847	1014	720	780	857
BOD$_5$ mg/l	329	-	901	-	387	398	-	504
pH	12.1	12.4	12.4	11.9	11.7	11.5	11.6	11.9

Sludge Cake Characteristics

Sludge cake is the discharge from the pressure filter and is about 58 inches diameter, 1½ inches thick, weighs about 200 pounds. Each pressure filter produces 83 cakes per cycle which are discharged to a storage bunker. Shear cables are positioned across the top of the bunker to break these large cakes into smaller pieces. Cake is conveyed through a series of drag conveyors and elevators to the incinerator. At normal operation cake moisture content is in the range of 36-38 percent and the appearance is dense, dry and textured. Extreme care should be exercised in selecting proper materials handling equipment to transport this cake to disposal. It is essential that minimum transfer and handling of the cake be performed from the filter to ultimate disposal. Excessive shear, breaking, dropping, grinding, or any form of handling results in plasticizing the cake to the consistency of dense putty or clay-like material. At these properties the plasticized cake material tends to be extremely sticky even at 62-64 percent solids and causes many problems due to poor release from drag and screw conveyors. The crumbly, dry, textured character of the original sludge cake can be maintained with minimum transfer and handling. Grinders were originally installed at Cedar Rapids in the incinerator feed system, however, these have been removed and are undesirable with this type of cake. By the time the cake has traveled through the materials handling system to the incinerator it has broken into small pieces of ¼ inch to 2 inch size. These pieces are generally in the form of balls due to the rolling action of the incinerator feed screw conveyor. Balls probably would not develop if this final conveyor were of a different design. However, the shape of the feed particles is not significant to the incinerator performance.

The specific weight of the discharged filter cake varies from about 107 to 114 pounds per cubic foot dependent upon moisture content and ash ratio. Under normal operating conditions the moisture content of the sludge cake varies only a few percentage points, therefore, the greatest influence on the specific weight of the cake is the ash content (ash/sludge ratio). The specific gravity of organic ash is in the range of 2.6 to 2.8. Laboratory determinations of specific gravity using water medium was 2.63; using hexane medium, was 2.78.

Cake discharged from the filter to the cake storage bunker is sheared by a series of cables which tends to bulk the broken cake. However, the impact of additional cake falling on the pile of previously discharged cake tends to cause recompaction due to the plastic properties of the sludge cake. Bulked filter cake was determined to be about 47 pounds per cubic foot, but this was for cake not exposed to impact and compressive forces due to dropping and stacking. That low figure might be valid for some conditions of limited cake storage where bulking is the major influence. Cake discharged from the bottom of the hopper under compressed conditions of superimposed loads of 10 to 12 feet depth had a specific weight of 83 pounds per cubic foot. This was determined by weighing many fixed volume samples in the bunker conveyor.

Conditioned sludge applied to the pressure filter forms a cake structure which almost immediately serves as the principal filter media. The filter cloth has served only as a base structure for this development, and after the cake has formed to 1/16 or 1/8 inch thickness, the influence of the filter cloth is negligible to the continued filter performance. As the cake continues to form and the filter void is filled, the total resistance increases rapidly developing full pressure differential in possibly 30-40 minutes as shown by a typical filter pressure chart, Figure 57.

Typical Filter Pressure Chart

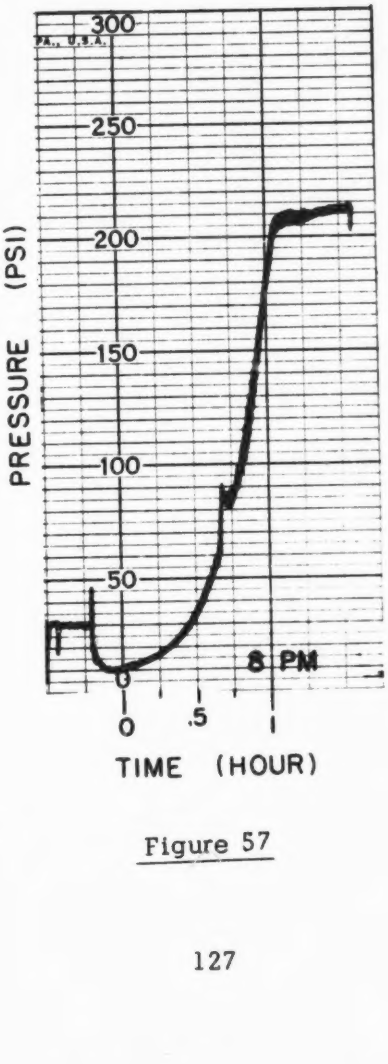


Figure 57

Continued full pressure differential is desirable to assure that most of the free water has the time to travel through the cake and underdrainage system to be disposed of as filtrate. The noteworthy point after the development of the full cake formation is that prolonged high pressure does very little to further dewater the cake. The moisture content may range only a few percentage points over a relatively wide span of sludge feed densities; and the cake moisture content is even less influenced by prolonged dewatering time. An example of this is found in Figure 58 where sludge cake densities are shown for different percent feed solids and cycle time. The curve is a best fit line using arithmetic probability points shown as dark points of average solids groupings.

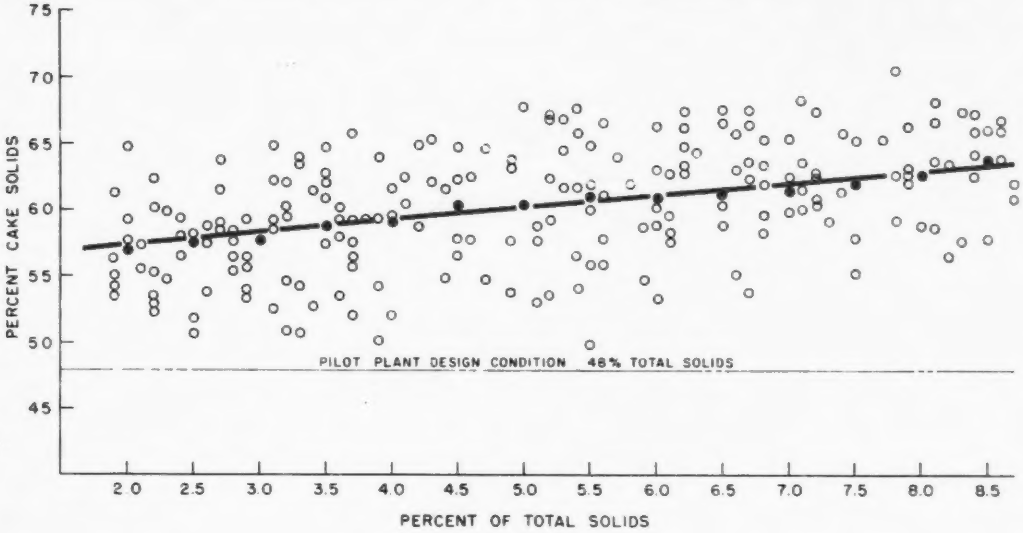


Figure 58

Grease and nitrogen determinations were made on composite samples for each shift of operation. This consisted of three composite samples per day made up from each filter run per shift. The average monthly analysis is as follows:

	April	May	June	July	August
Nitrogen mg/l	1.83	1.58	1.16	0.86	0.87
Grease mg/l	2.40		1.36	1.80	1.70

Phosphate Removal

There has been increasing emphasis upon the removal of phosphorous from waste. It then becomes very important that the dewatering of phosphate rich sludges should produce a filtrate low in phosphorous. Phosphorous generally is removed by precipitation using three commonly available cations, Al^{+3}, Fe^{+2}, or Fe^{+3}, and Ca^{+2}.

In the process of sludge dewatering Fe^{+3} and Ca^{+2} are present as coagulants. Presumably because of the negative oxidation reduction potential of the sludge a portion of the iron is reduced to Fe^{+2}. Consequently three precipitates of phosphorous could be possible; ferrous phosphate ($Fe_3(PO_4)_2$), ferric phosphate ($FePO_4$), and calcium phosphate or more properly calcium hydroxyapatite ($Ca_5(OH)(PO_4)_3$).

Phosphorous removal through the pressure filter was investigated by using the resistance meter on the bench and spot checking actual phosphorous removal through the full scale filter.

Bench Scale Studies

Since most of the data from the pressure filter was obtained when the sludge was conditioned with lime and ferric chloride, some bench work was also performed using the resistance meter and conditioning only with ash. In this test a portion of the same sample was conditioned with fly ash using ash/sludge ratios of 0, 1.0, 1.5, 2.0, 2.5 and 3.0. Separate portions were also conditioned with sludge ash using the same ash/sludge ratios. The samples were then filtered through the resistance meter and the filtrates were analyzed for total phosphate. The results are graphically displayed in Figure 59 and listed in Table X.

Table X

	Fly Ash			Sludge Ash		
A/S Ratio	Phosphate as P	Percent Removed	pH	Phosphate as P	Percent Removed	pH
0(no ash)	25.6 mg/l			25.6 mg/l		
1.0	12.3	51.9	8.1	12.3	51.9	8.3
1.5	10.4	59.4	8.4	10.4	59.4	8.3
2.0	10.0	60.9	8.4	8.3	67.6	8.5
2.5	8.6	66.4	8.5	6.8	73.4	8.6
3.0	7.6	70.3	8.6	6.2	75.8	8.7

The removal of phosphate was nearly identical for samples treated with fly ash and the samples treated with sludge ash.

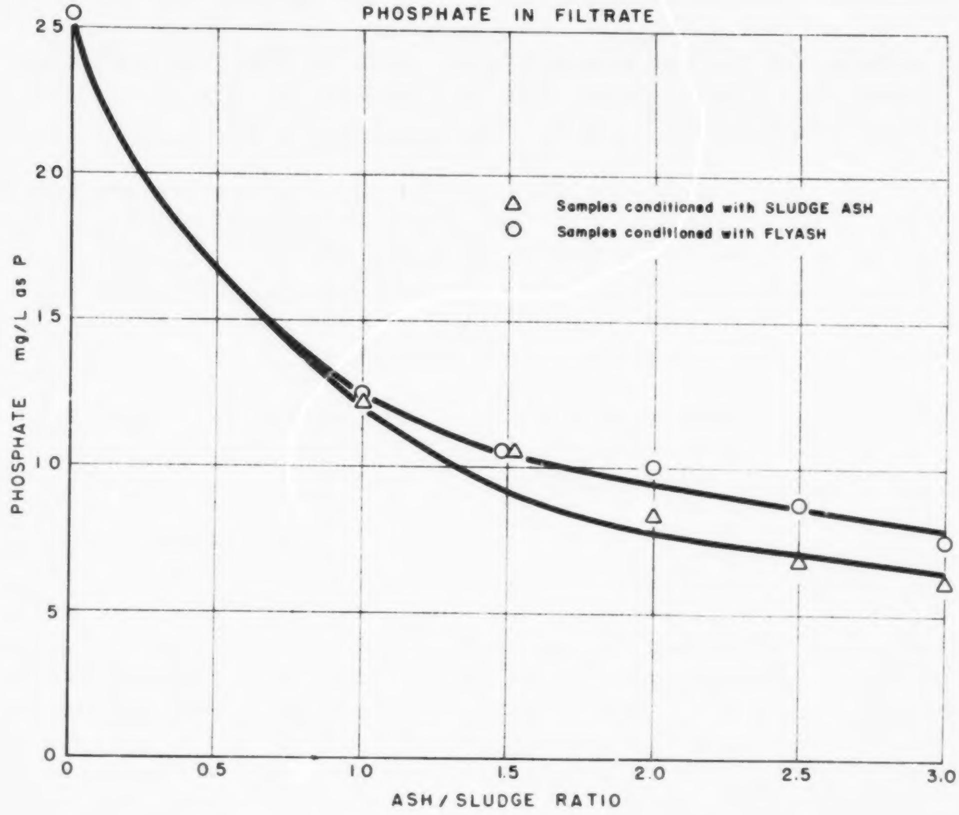


Figure 59

Full Scale Studies

During the course of the study several runs on the full scale filter were made using only fly ash and no chemicals (ferric chloride and lime).

The phosphate levels in the raw sludge and the removals are:

Run No.	Total Phosphorous as P	Soluble Phosphorous as P	Percent Removed
1 Sludge	516	100.5	91.3
Filtrate	8.75	2.5	
2 Sludge	528	73.5	74.8
Filtrate	18.5	4.5	

The percent phosphorous removal was based upon the soluble phosphorous in the sludge and total phosphorous remaining in the filtrate. Presumably the insoluble phosphorous in the sludge is principally discharged with the sludge cake.

$$\text{Percent Removal} = \frac{\text{Soluble P Sludge} - \text{Total P Filtrate}}{\text{Soluble P Sludge}} = \frac{100.5-8.75}{100.5} = 91.3\%$$

Additional data on phosphorous removal was obtained when ferric chloride and lime were also used for conditioning along with ash. This data is presented in Table XI.

Table XI

Phosphate Removals by Pressure Filtration
Using Lime, Ferric Chloride and Sludge Ash as Coagulants

Date	Sludge Unconditioned		Sludge Conditioned		Filtrate	
	Total	Soluble	Total	Soluble	Total	Soluble
7-20-72	500 mg/l	*	1204	*	2.4	*
7-24-72	592	*	1596	*	16.8	*
7-25-72	564	*	1448	*	16.8	*
7-26-72	786	*	1752	*	11.8	*
7-27-72	520	*	1124	*	16.8	*
7-27-72	546	*	1256	*	14.8	*
7-27-72	514	*	1528	*	19.8	*
8- 3-72	442	96	1732	96	13.2	< 0.1
8- 7-72	980	91.5	1348	131	12.3	< 0.1
8- 9-72	446	104	1656	134	8.75	*
8-10-72	1150	142	2890	172	3.65	*

* Not determined

The percentage reduction of phosphorous realistically should be based upon the soluble phosphorous in the unconditioned sludge versus the total phosphorous in the filtrate. This will give a somewhat lower percentage of removal than a percentage removal based upon total phosphate in the conditioned sludge versus total phosphate in the filtrate.

Therefore, the comparison of the two processes of phosphate removal, ash only conditioning versus ash plus chemical conditioning, may be difficult unless the basis of comparison is the same. That is, the comparison of total and soluble phosphate.

Percentage removals of phosphorous through the pressure filter averaged 86.7 with the soluble phosphorous in the sludge averaging 108.4 mg/l, and the filtrate total phosphorous averaging 14.4 mg/l. On a total phosphate basis the percent removal was 97.8.

Filtration with Fly Ash

During the course of the study a program was developed to operate the full scale pressure filters using only fly ash as a digested sludge conditioner. The limited work that was done affirmed the fact that fly ash alone (without chemicals) was sufficient to dewater digested sludge. This also confirmed the conclusions of the original pilot plant report wherein only fly ash was used as a sludge conditioner.

However, the use of fly ash posed many problems in materials handling and conveying. Fly ash tends to fluidize and with the ash handling equipment used at Cedar Rapids the accurate control of admix quantities was impossible. Our experience would indicate that gravimetric feeders are not adequate for feeding materials which fluidize.

Another problem which occurred was conveying the filter cake. Although the percent cake solids averaged 55.5 percent for all runs, the fly ash cakes were virtually impossible to convey. After being conveyed from the bottom of the bunker the cake turned into a sticky mass which refused to fall off the cake elevator flights. Therefore, only limited work was performed using the full scale filter.

Below is the summary of the best filter runs using fly ash as a conditioning agent.

Percent Solids	= 5 percent
Ash/Sludge	2.57
Cycle Time	1.80 hours
Yield	0.57
No. Run	4

Other runs which were made at lower percent solids in the feed sludge were dosed with extreme amounts of ash (A/S = 4.5 to 6.2) because of the fluidizing of the fly ash.

Filtrate

Filtrate from dewatering, using fly ash, showed great differences from the filtrate using sludge ash and chemicals. Not enough runs were made to draw any generalizations.

Following is a comparison of filtrate from one day of operation.

		Fly Ash Conditioned	Sludge Ash and Chemicals Conditioned
pH		9.3	10.4
Suspended Solids	mg/l	148	43
Color	Units	50	150
COD	mg/l	480	830
BOD_5	mg/l	250	470

It can be seen that the filtrate appears to have a lower COD, BOD and color when the sludge is conditioned with fly ash rather than sludge ash and chemicals.

Dewatering of Raw Primary Sludge

The filterability of raw sludge as compared to digested sludge at Cedar Rapids is generally much better. Specific resistance values for raw sludge range from values of 120 to 300 x $10^{12}cm^{-2}$, with an average value of approximately 150 x $10^{12}cm^{-2}$.

Because of better filterability of raw sludge, a program was set up to develop chemical and ash dosage requirements so that raw sludge could be run on the full scale filter.

Bench testing and testing using the six inch pilot pressure filter indicated that satisfactory performance could be obtained using the chemical dosages for digested sludge and using a 1/1 ash/sludge ratio at all solids concentrations.

Accordingly, in-plant arrangements were made to empty one storage tank and to pump raw sludge from a gravity thickener to the empty storage tank. First filter runs with raw sludge were made November 28, 1972, and during the month of December over 60 filter runs were made on raw sludge. The runs were generally quite satisfactory.

During the month of December, the raw sludge pumped to the sludge holding tanks averaged 3.7 percent solids. During this same period, sludge from the holding tanks to the process averaged 5.78 percent solids, and much clear supernatant could be decanted from the holding tanks.

Table XII

Summary of Raw Sludge Filtration Data

5%			
Average Ash/Sludge		1.20 range	1.56 to 1.00
Chemical Cost	$/ton	6.26 range	$7.01 to 4.67
Yield	lb./sq.ft./hr.	0.68 range	0.97 to 0.30
Cycle Time	hrs.	2.03 range	1.33 to 3.75
Number of runs - 27			
6%			
Average Ash/Sludge		1.19 range	1.44 to 0.68
Chemical Cost	$/ton	5.17 range	$5.61 to 4.71
Yield	lb./sq.ft./hr.	0.90 range	1.41 to 0.29
Cycle Time	hrs.	1.75 range	0.83 to 3.83
Number of runs - 22			
7%			
Average Ash/Sludge		1.13 range	1.17 to 0.91
Chemical Cost	$/ton	4.97 range	$5.29 to 4.05
Yield	lbs./sq.ft./hr.	0.92 range	1.30 to 0.39
Cycle Time	hrs.	1.69 range	1.08 to 3.25
Number of runs - 8			

In Table XII, the results for the filtration of raw sludge are given. Data are given for 5, 6 and 7 percent solids. Other data for 3, 4, 8 and 9 percent solids are not given. These were insufficient filter cycles to interpret a curve. As can be noted, yield, cycle time and chemical costs all improved as the percent feed solids increased from 5 to 7 percent.

Yield versus cycle time is shown in Figure 60 for raw sludge. Yields are all quite good compared to digested sludge, basically due to low ash ratios of approximately 1.2.

Filter Cake

Pressure filter cake averaged 54, 56 and 58 percent solids for filter feed solids concentrations of 5, 6 and 7 percent solids respectively. Volatile solids content of the cake was higher than for digested sludge, averaging 25, 27 and 29 percent for 5, 6 and 7 percent feed solids respectively. These values compare to 11-13 percent volatile solids at comparable dewatering conditions with digested solids. These levels of volatility were much higher than that necessary (approximately 11 percent) to provide autocombustion in the incinerator. No difficulty was experienced in conveying the raw sludge filter cake.

Filtrate

The most significant difference in filtrate quality caused by filtration of raw sludge was the very high BOD's and COD's of the filtrate. In Table XIII is given comparative values of filtrate COD's and BOD's for digested sludge and raw sludge during the two month period of November and December 1972. These high BOD's are the result of the formation of soluble volatile acids in the sludge. These volatile acids depressed the pH of the raw sludge to values of approximately 5.0 (± 0.2). The filtrate pH averaged 11.6 due to sludge conditioning with lime.

Filtrate suspended solids showed no basic difference than the filtrate from digested sludge. Filtrate samples from all runs averaged 79 mg/l. During this same period suspended solids from the digested sludge filtrate averaged 90 mg/l. As shown in Table XIII there was no significant difference between filtrates.

Table XIII

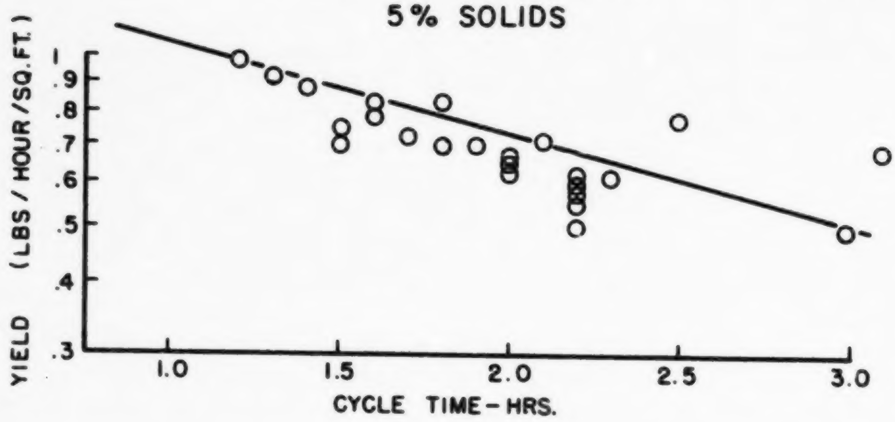
Comparative Values Dewatering Raw and Digested Sludge

	COD mg/l		BOD_5 mg/l		BOD_5/COD (%)	
Sludge	Digested	Raw	Digested	Raw	Digested	Raw
Maximum	1080	13,500	740	10,300	70	94
Minimum	270	2,000	140	1,700	41	72
Mean	510	7,080	300	5,700	56	81
Median	470	6,600	250	5,200	58	81

Comparative Values of Suspended Solids in Filtrate

Type of Sludge	Maximum	Minimum	Mean	Median
Raw	430 mg/l	22 mg/l	79 mg/l	53 mg/l
Digested	345 mg/l	20 mg/l	90 mg/l	61 mg/l

RAW PRIMARY SLUDGE
5% SOLIDS



RAW PRIMARY SLUDGE
6% SOLIDS

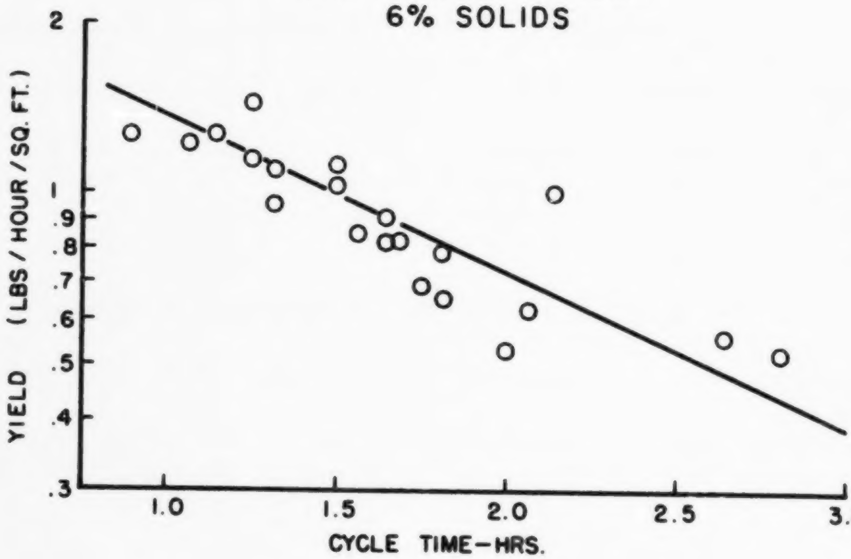


FIGURE 60

SECTION X

PROCESS EVALUATION

Economic Evaluation

Basis for Economic Evaluation

The labor requirements were determined for sludge dewatering and sludge incineration for both operation and maintenance. These labor costs were based upon actual wages paid for 1972 including an allowance for fringe benefits such as insurance, vacation, workmen's compensation, retirement, etc. Hourly rates were \$4.58 for the operator and maintenance shift foremen and \$4.17 for assistant. Management or laboratory costs are not included.

The sludge dewatering plant was bid in December 1968. To be consistent throughout this report, all costs have been interpreted to 1972 costs. Earlier in this report the pilot plant evaluation costs predictions made in 1968 were interpreted to current 1972 costs, therefore the capital costs for the construction of the sludge dewatering facility for which bids were received in December 1968 have also been interpreted to 1972 using the Engineering News-Record index of cost trends. The equipment costs increase for that period has been interpreted to be 20 percent. The capital costs used in this evaluation are the actual bid costs for December 1968 scaled upward as outlined to reflect what these costs might have been for 1972.

Chemical costs are those actually paid for pebble lime and liquid ferric chloride. Power costs were assumed to be \$0.02/KWH for filtration although it may actually be less due to a favorable power demand-power consumed ratio. Power consumed for each filter is separately metered, however, total power for the dewatering and incineration processes is also metered and is the basis for determining power company billings. High power demand charges for the incinerator high energy off-gas scrubber distort the power billings and make it necessary to interpret the filtration power charge as indicated above. Power consumed for filtration was measured by the separate watt-hour meters during observed filtration cycles and was determined to be 40 KWH per hour of actual filtration.

Cedar Rapids has two pressure filters and all labor costs are determined for the operation of two units on a full time basis. All costs were reduced to dollars per filter hour operation so that the final cost would be evaluated to filtration capacity regardless of the operating capacities experienced during the study.

	Man Hours/Hr. Filtration		$/Hr. Filtration	
Operation Foreman	0.90		4.13	
Assistant	0.33	1.23	1.37	5.50
Maintenance Foreman	0.17		0.76	
Assistant	0.33	0.50	1.37	2.13
		1.73		$7.63

Labor costs per filter (÷ 2 units)		$3.82/hr. filtration
Power costs 40 KWH/hr. filtration x $0.02		0.80/hr. filtration
Chemical costs	ferric chloride ($130.00/ton)	0.065 /pound
	lime ($ 22.75/ton)	0.0114/pound

Capital costs 20 years at 4 $\frac{3}{4}$%
5$\frac{1}{2}$ days per week - 286 working days/year

Building $ 417,000
Equipment 1,255,000
$1,672,000 = $19.10/hr ÷ 2 units = $9.55/hr. filtration

No cost allowance has been made for fly ash haul to the plant or product ash from the plant. This cost may vary considerably from a small cost as in the case at Cedar Rapids, to a significant cost if the fly ash was purchased and the haul distance was great. The basic operation is with incinerated sludge cake ash recycled for sludge conditioning. However, as a part of this report and to follow up on the original pilot plant study wherein fly ash was used as a sludge conditioner, power plant fly ash was transported and used as a sludge conditioner. The fly ash was obtained from the same power generating plant as was used in the pilot plant study. The ash was obtained at no cost. Transportation was arranged with a contract hauler using a bulk cement transport trailer truck. Fly ash was loaded by gravity chute at the power plant, transported about three miles distant, unloaded by blowing into the ash storage bins. The cost for transporting fly ash filter aid was $4.00 per ash ton. The filter performance using fly ash filter aid averaged about $6 to $8 per ton dry sewage solids as the cost of hauling fly ash to the process. Product ash haul to disposal would vary considerably with local conditions and would be less significant with organic ash than with fly ash filter aid. At Cedar Rapids it was possible to dispose of the excess ash on the site, therefore haul-out costs were not accounted.

Process Costs

The total cost for pressure filtration is composed of four separate costs. These costs are labor for maintenance and operation, power, chemicals for conditioning, and capital investment.

Labor costs were determined, as described before, to be \$3.82 per hour per filter. Power costs were determined in the same manner to be \$0.80 per hour per filter. Likewise capital costs were determined to be \$9.55 per hour. In the compilation of the data each cost was determined as follows:

$$\text{Cost/ton} = \frac{\text{cycle time (hours) x cost factor}}{\text{tons dry solids/filter}}$$

Cycle time represents the time to complete a filter run. Added to filtration time, approximately 0.4 hour is required to discharge the cake, refill the filter with filtrate, and precoat, and is referred to as turn around time.

$$\text{Cycle time (hours)} = \text{filtration time (hours)} + 0.4 \text{ hours turn around}$$

The cost factor represents the dollar per hour value placed upon each cost.

Chemical costs were determined for each cycle by measurement of actual quantities of lime and ferric chloride used to condition the sludge.

Figures 61, 62 and 63 show cost data for $4\frac{1}{2}\%$, $5\frac{1}{2}\%$, and $6\frac{1}{2}\%$ solids respectively. Only operating costs, which includes both power and labor, capital costs and total costs are shown.

These three cost curves were determined by using costs for each run which were determined by a computer, then taking these values and performing a linear regression analysis based upon the relationship log cost versus log yield.

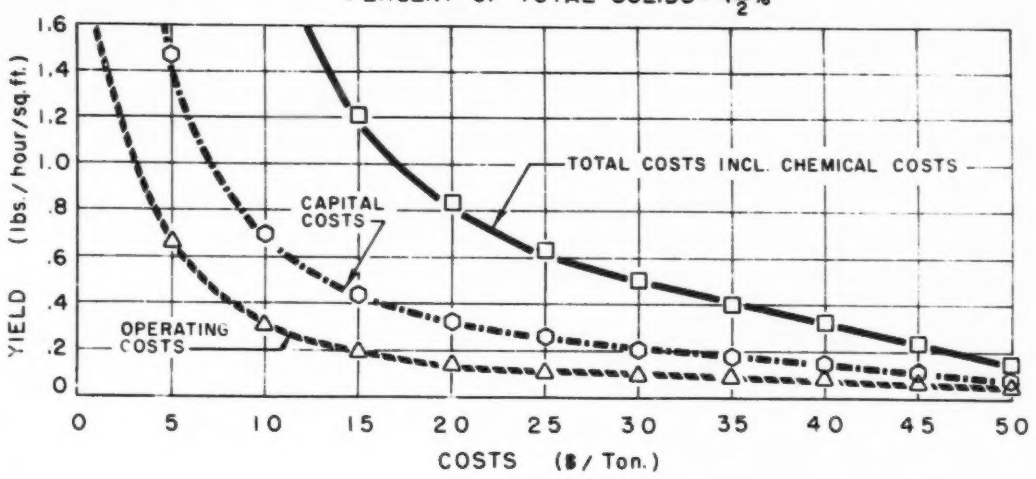


Figure 61

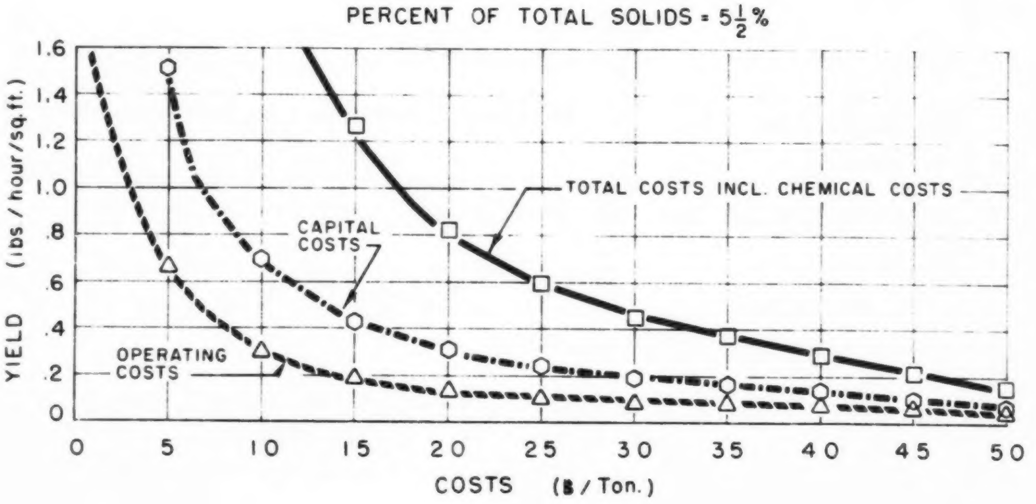


Figure 62

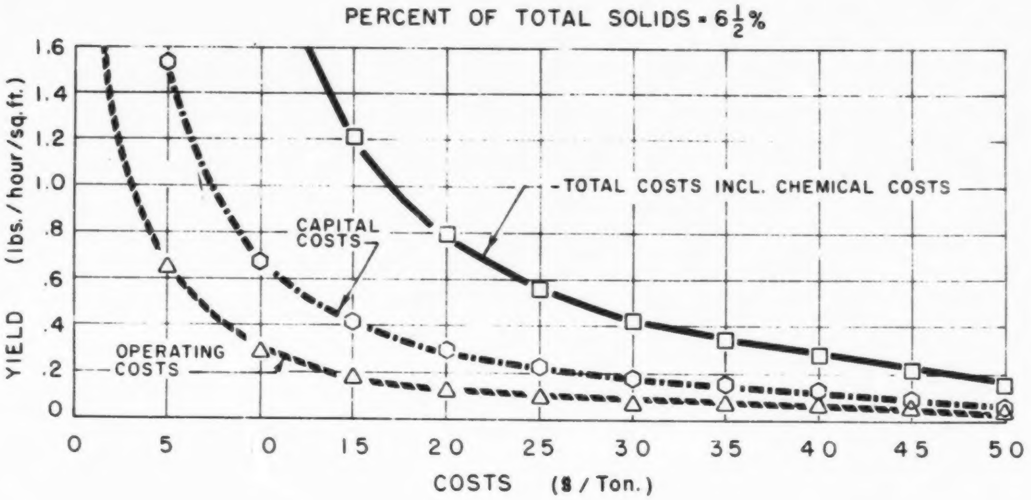


Figure 63

The following correlation coefficients were obtained.

Percent Solids	Operating Cost	Capital	Total
4½	-0.985	-0.985	-0.875
5½	-0.991	-0.991	-0.925
6½	-0.988	-0.988	-0.865

The range of cost experience is shown in Table XIV.

Table XIV
Process Costs

	Maximum	Minimum	Average
4½%			
Operating	11.89	2.84	5.83
Capital	24.05	5.87	12.05
Total (including chemical)	43.64	17.72	26.83
5½%			
Operating	9.32	2.65	4.69
Capital	19.27	4.54	9.71
Total	34.05	14.80	21.69
6½%			
Operating	6.85	1.47	3.83
Capital	14.17	3.04	7.91
Total	28.44	11.22	18.20

Negative correlation coefficients mean an inverse correlation. The correlation coefficient can have values from -1 to +1 where an absolute value of 1 means a perfect correlation whether the correlation be direct (+1) or inverse (-1). A value of zero indicates no correlation or simply a random scattering of points.

All correlation coefficients shown above are negative and closely approach 1 in absolute value. As can be expected, costs related to hours operation decrease as productivity or yield increase, and as can be seen from the graph, these costs decrease in a predictable manner.

The correlation coefficient for total costs are lower because of the seemingly lack of correlation of chemical dosages (costs) with yield. Correlation coefficients for chemical costs were developed in the same manner as the other costs. The correlation coefficients were 0.122, 0.217 and 0.005 for $4\frac{1}{2}$, $5\frac{1}{2}$ and $6\frac{1}{2}$ percent solids respectively. These are very low and poor correlation coefficients.

Chemical costs showed poor correlation for several reasons, they are:

1. Varying specific resistance of digested sludge.
2. Lack of sufficient conditioned sludge pumping capacity to the filter.
3. Over-ranging of chemical feeders.

The first two reasons are interrelated. As has been brought out in previous sections, chemical (lime and ferric chloride) dosages were set up to properly condition sludge with high resistance. Since the sludge varied considerably in resistance, some sludge was over-conditioned, While over-conditioning the sludge should have resulted in a higher yield by giving a higher filtration rate, the maximum filtration rate of the filter is mechanically limited by the capacity of the two filter feed pumps. Later it was demonstrated that using all four filter feed pumps for one filter significantly decreased filtration time (up to 50%), by meeting the true capacity of the filter for sludge. Because of this mechanical limit, chemical conditioning was only partially effective in raising yields.

The other problem which occurred was the oversizing of the chemical feeders specifically the ferric chloride feeder. Because of this oversizing, the feeder operated in the lower 20 percent of its capacity where accuracy was at a minimum.

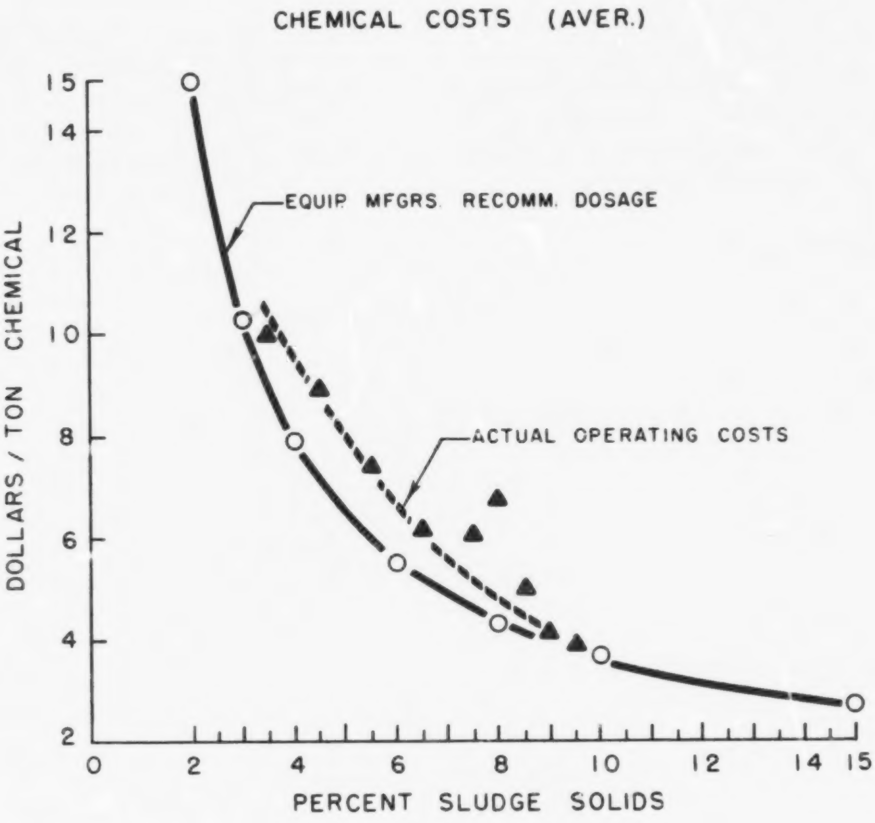


Figure 64

Figure 64 shows average chemical costs per ton of dry solids versus percent solids. Also shown is the chemical cost curve for the proposed costs developed from work on the resistance meter by Dr. O. Busse. As can be seen from the graphs, operating costs were somewhat higher than theoretical yet followed the cost curve very closely.

Optimization of Costs

From the figures present it becomes quite apparent that real cost savings can be made by pre-thickening of sludge solids. Thickening the incoming sludge from 3 to 6 percent decreases chemical costs at Cedar Rapids from \$10.00 per ton dewatered to \$6.00 per ton dewatered. Ash requirements also decrease from 3 pounds per pound of sludge solids to 1½ pounds per pound of dry solids. See Figure 45. Obviously the less volume of the filter taken up by ash means more capacity for sludge. Cost advantages for pre-thickening decrease for influent solids in excess of 6 percent.

All costs developed in this section, especially those for chemicals, are specifically for Cedar Rapids and do not have general application, and do not form a sound basis of comparison with other processes. Reference is made to the previous section of this report wherein chemical costs without ash on a vacuum filter at Cedar Rapids could not be lowered below $20.00 per ton because of the high specific resistance of the sludge.

Process Evaluation

Having operated the pressure filter pilot plant at an early date when no analytical data was available for evaluation and guidance, and having developed the design criteria for a full scale plant from those pilot studies, Cedar Rapids has observed the process performance without bias to other installation performances. As stated previously, the pilot plant performed reasonably well with good data correlation, however, the pilot process was hand controlled and should reasonably be expected to perform well. The full scale plant process is highly automated and therefore, reflected some of the malfunctions associated with equipment failure from numerous interrelated units.

Some unsatisfactory process performance was experienced at Cedar Rapids during the early days of start-up and check-out. Much of this was related to inadequate training of the operating personnel, particularly those of the equipment manufacturer. This is mentioned here because so many visitors to this facility during more than a year of start-up and check-out prior to acceptance were aware of this problem. It is understandable that the first major installation would offer some degree of challenge on start-up and that experience would make the next one easier.

Looking backward to the expectations of the pilot plant study, and to the process performance observed in the full scale plant, it is obvious that the pressure filter process is dependable, economical, and offers distinct advantages over other forms of sludge dewatering based upon Cedar Rapids sludge. The process has achieved a higher quality product with greater capacity in the full scale plant than was predicted by the pilot plant study. Cake quality is maintained over a broad range of sludge solids concentration, particularly at the low solids level (2.5-4.5%) which were not previously experienced in the pilot studies.

Incineration of Sludge Cake

The primary subject of this report is dewatering of digested sludge by pressure filtration. The process at Cedar Rapids is dependent upon recycled organic sludge ash as a sludge conditioner and filter aid. It would, therefore, seem appropriate to briefly review the incineration of the pressure filter cake and the recycle process. As previously described, cake is discharged from the pressure filter to a storage bunker and transported through a series of conveyors to the incinerator.

The cake incineration process starts with the bunker feed conveyor which is a variable speed drag conveyor. This speed control on the bunker discharge serves to regulate the rate of cake feed to the incinerator and is manually set by the operator. Automatic temperature controls regulate the auxiliary fuel feed, if required, to maintain hearth temperatures. The incinerator is a conventional multiple hearth furnace previously described.

When pressure filter performance is maintained close to design ash/sludge ratios of 1.5/1, it has been observed that furnace performance can be controlled so that the cake feed is autocombustible with 11 to 13 percent volatile solids content. Under normal filter operation the moisture content of the cake falls in a narrow range of values regardless of the raw sludge feed density, therefore, the major factor in controlling the percentage of volatile solids in the cake is the ash/sludge ratio. As the ash/sludge ratio increases, the volatile content of the cake decreases.

Numerous heat balances have been programmed on the furnace operation, however, these are not the subject of this report and it will suffice to summarize that the filter cake from digested sludge will burn without auxiliary fuel when furnace feed conditions are held within a reasonable range (± 50%) of the design ash/sludge ratio and cake feed rate.

The internal conditions of the furnace are significantly different while burning cake with recycle ash. The internal atmosphere is extremely dusty even under minimum excess air control and good balance. A considerable amount of ash is picked up in the gas stream in the upper drying hearths and carried with the off-gas. This added particulate volume makes an adequate scrubber system even more important. Cedar Rapids' incinerator off-gases are directed to a wet venturi high energy scrubber operating [illegible] about 30 inch water column pressure drop. Scrubber water is returned to the treatment plant and contains 4,000 - 10,000 mg/l suspended solids. This seemingly large quantity of fine ash carried out in the scrubber water was of considerable concern for some time. The general philosophy for successful pressure filtration with ash filter aid was that the fine ash particles were extremely important to developing a good dewatering cake. The Germans had experienced and reported this requirement. In the early periods of check-out the apparent shortage of finer ash in the recycled conditioning ash was considered a factor of poor filter performance. After considerable bench work with and without the fine ash carried out in the scrubber water it was determined that the absence of this fine ash did not influence sludge conditioning and filterability.

Incinerator ash passes through a hammer mill grinder which originally contained 1/16 inch screens. These were later enlarged and ultimately increased in size to $\frac{3}{4}$ inch. It has been demonstrated at Cedar Rapids that, contrary to the European practices, grinding recycled ash is not necessary, and for ordinary protection of the process equipment simple screening for foreign and oversized material would be satisfactory.

Considerable difficulty has been encountered with incinerated ash in storage and feeding equipment. Both organic ash and fly ash readily fluidizes and compacts over a narrow range of external influence. Elevated ash storage bins were equipped with customary bin vibrators to assure even flow to the outlet feeders. It was necessary to employ a trial and error adjustment of both frequency and magnitude of vibration to get the correct adjustment of the external influence to assure satisfactory flow of ash to the feeder. Until the correct combination of magnitude of impact and frequency was determined it was unpredictable whether the output ash from the storage bin to the feeder would fluidize or compact. At times it would fluidize at one set of conditions and at other times it would compact. Temperature of ash leaving the incinerator is normally about 250° F. and this causes no problems in the elevated storage bins. However, due to process upsets the ash temperature may go to 500-700 degrees F. for short periods. These high ash temperatures into a partially filled bin apparently caused a temperature turbulence and a fluidized condition which would progressively work through the ash in storage to the outlet and thus upset the feeder performance.

Proper precoat is essential to satisfactory filter performance and precoat pressure is an important control for assuring good cake formation. Normal precoat pressure for a clean filter media starts at about 25 psig and ranges upward to 40 psig. Above this operating range it appears that gradual deterioration of filter performance often occurs. To operate within the precoat pressure range it is necessary to have properly conditioned raw sludge applied to the filter and to have had the proper quantity of ash precoat.

Improperly conditioned raw sludge will cause some buildup on the media and thus gradually raise the successive precoat pressures. Sometimes this problem can be corrected by dropping the quantity of precoat from 150 to 50 pounds thus forming a new cleavage plane encouraging the discharged cake to remove some accumulated precoat. Improper quantities of precoat may cause a gradual buildup whereby the excess ash creates a shear plane distant from the filter media so that when the cake breaks away excess precoat remains on the media.

An operating experience which has reduced the precoat pressure, or at least prolonged the need for a complete filter wash is that of extending filter fill cycles. During the fill cycle filtrate water is pumped into the filter and by continuing recirculation of the overflow it is sometimes possible to perform some degree of washing. This practice is not totally successful, but it may reduce the precoat pressure by 10 psig which may extend the need for a complete filter wash for a few days.

Estimated filter runs between media wash is about 100 runs, or 150 to 200 filter hours of operation. The method of washing a pressure filter could stand much improvement. The wash rod with a series of high pressure nozzles originally furnished is no longer used. A more successful system has been a single high pressure nozzle with a broad discharge operating at about 750 pounds pressure. Commercial grade detergent is used in the washing solution.

Equipment Evaluation

From the beginning of check-out considerable equipment difficulties were experienced with the full scale plant. These are equipment problems beyond those normally expected or experienced during start-up and check-out. Many of these unusual problems related to poor workmanship and fabrication details of the pressure filter and accessories equipment. These problems were greatly compounded by manufacturers' personnel inexperienced in detailed process and equipment performance. Cedar Rapids became a training center for these people which accounts for the fact that it took almost 18 months to approach a reasonable performance. Numerous modifications to the mechanical features of the equipment have been made, along with the electrical and process controls. Some of the poor fabrication quality control, such as poor machining and assembly, resulted in failures and were subsequently repaired or replaced with improved equipment.

Some of the major problems experienced, or which are still of concern are mentioned here for purposes of objective reporting. It is not the intent to criticize the equipment supplier at this late date, because in fact, periodic improvements are still being made to the equipment at Cedar Rapids by Beloit-Passavant as of January 1973. The equipment supplier has continued to respond to the problems as they have occurred and we have every reason to think that this cooperation will continue.

Pressure filter plate warpage has been a major problem from day one. Some of this has been attributed to the early days of operation when the plates were inadequately shimmed, however, adequate shimming of the stay bosses has not totally eliminated this problem. Warpage occurring in the plate diaphragm transfers bending to the plate frame which in turn accelerates plate gasket deterioration due to warped gasket seating plane. Plate alignment also affects gasket life wherein poor alignment, or repositioning of the plates each cycle, causes a reshaping of the plate gasket leading to premature failure. Plate gasket life is short from whatever the cause, and is still unsatisfactory in our opinion.

Cake and incinerated ash grinders were originally installed because the German process recommendations stressed this requirement. However, experience at Cedar Rapids confirmed that cake grinding was undesirable and that incinerated ash grinding was unnecessary. Filter cake broke and crumbled sufficiently in the conveying system for delivery to the incinerator. Cake passing through the grinder tended to knead to a plastic like ball and under high moisture content of about 55 percent cake into the grinder came out more as an extrudable paste. The cake grinder screen was first removed, then the rotor as well, so that in effect no grinder exists. The incinerated ash grinder was originally a 1/16 inch screen. This fine screen had a tendency to clog even though the ash particles were much smaller. At present the screen is $\frac{3}{4}$ inch and serves as a trap metal safety device with the rotor operating to assure through-put. No measurable effect has been detected in ash particle size.

Sludge nuclear density meters are of considerable value to this process to have an immediate sludge density reading to guide the operator in setting ash feed for proper ash/sludge ratio. The density meters originally installed did not hold calibration. This seems to be the major complaint of most sludge nuclear density meters. New meters are on order, and some calibration techniques developed in the field by plant personnel will be used which have proven more dependable than those practiced by the meter manufacturers.

Ash feeders used at Cedar Rapids are of the gravimetric type. These were installed for accurate proportioning of ash. Experience in handling both fly ash and organic ash would suggest that this refinement is not necessary and that volumetric feeders would be satisfactory and probably more trouble free. Feeder problems are associated with the unpredictable bulking and compaction characteristics in storage. It is difficult to cover both extreme properties of ash in the performance range of the storage bin activators and ash feeders.

Ferric chloride feeders were very inconsistent throughout the program, however, this was primarily due to improper sizing and not really a fault of the equipment. The feeders were oversized and operated on the low range where repeatability was poor.

Materials handling equipment, particularly filter cake conveyors were the cause of many problems. The primary cause of nearly all conveyor problems was the fact that no experience had existed in discharging filter cake to bunker storage. The bulk density of broken filter cake was greatly underestimated at about 48 pounds per cubic foot. This is reasonable for filter cake as it is sheared and bulked. However, in bunker storage the cake recompacted dropping into the bunker so that the bulk density was about 83 pounds per cubic foot at the drag conveyor discharge.

Filter feed pumps are hydraulic driven ram pumps having a variable capacity, variable head characteristic. Pumps having these characteristics are desirable as the filter cycle pressure develops and the input diminishes, however, a more suitable primary pump should be provided to meet the early demands of the filter, particularly on a filter installation as large as Cedar Rapids. Considerable experimentation was carried out by the plant operating personnel wherein filter performance was greatly improved by placing all four filter feed pumps on one filter rather than the normal two units. Prolonged slow feed rates to the pressure filter to form satisfactory cake development is uncertain from our observations with sewage sludges and ash precoat. Where a precoat system is not used, it may be desirable and even necessary to slowly develop a cake on the cloth to provide a protective zone so that the solids are not driven into the cloth upon increasing pressure. With a precoat system the protective zone is already established and through-put should be as rapid as possible. Increased filter supply could also be accomplished by increased equalization tank capacity, but in that system some problems of maintaining air volume, level controls, solids suspension, and mechanical controls offset the advantages of additional direct pumpage to the filter.

The mix tanks have mechanically failed in many ways from broken shafts, bearing failures, gear misalignment, stuffing boxes, blades falling off and others. Many of these failures have been attributed to poor shop workmanship and probably could happen to any manufactured product at some time. Modifications and repairs have been made and are continuing and hopefully will be eliminated with better design and fabrication.

The pressure filter has been in service in this country for fifty plus years in industrial processes. It has only been in recent years that its application to sewage sludge dewatering has been seriously considered in Europe and more recently in the United States.

The pressure filter is nearly the same age as mechanical type waste water treatment yet it has taken almost as many years to develop a serious interest to its application to sewage sludge dewatering.

Equipment and process improve with experience of people in specific applications, therefore, it is reasonable to have confidence that the pressure filter application to sewage sludge dewatering will be improved with time and that specific experience.

At Cedar Rapids the pressure filter process has performed well, with many of its problems solved by the human beings who created those problems in the first place.

SECTION XI

ACKNOWLEDGMENTS

The participation and contributions of Mr. George W. Milligan, P.E., Water Pollution Control Plant Engineer, and Mr. Sabry Kamhawy, Chief Chemist, are particularly noted and were vital to the preparation of this report.

A special thank you is extended to the entire staff of plant operators, chemists, and maintenance personnel who contributed and cooperated in the success of this program.

The Honorable Mayor and City Council have been active in this program from the preliminary planning through the final report as has the Director of Public Services. These dedicated and concerned people authorized the program which otherwise would never have been undertaken.

Donald J. Canney, P.E., Mayor

Commissioners
John D. Oberthien
Richard L. Phillips
Stanislavs Reinis
Harold G. Schaefer
Richard H. Jensen, P.E., Director of Public Services

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SECTION XII

GLOSSARY OF TERMS

A/S	Ash to Sludge Ratio
BOD_5	5 Day Biochemical Oxygen Demand
CC (cc)	Cubic Centimeter
CFM (cfm)	Cubic Feet per Minute
COD	Chemical Oxygen Demand
Cu. Ft.	Cubic Feet
Ft.	Feet
Gals/Min/Sq.Ft.	Gallons per Minute per Square Foot
Gals/SF/Hr.	Gallons per Square Foot per Hour
In.	Inches
In. W.C.	Inches Water Column
KWH	Kilowatt Hour
Lbs. (or #)	Pounds
Lbs./Cu. Ft.	Pounds per Cubic Foot
Lbs./SF/Hr.	Pounds per Square Foot per Hour
μ	Micron
mg/l	Milligrams per Liter
ml	Milliliter
mm	Millimeter
MPR	Minutes per Revolution
%	Percent
pH	Hydrogen Ion Concentration
psig	Pounds per Square Inch Gauge
r	Correlation Coefficient
r	Specific Filter Resistance ($10^{12} cm^{-2}$)
Rev.	Revolution
S.S.	Suspended Solids
Sq. Ft.(S.F.)	Square Feet
T.F.	Trickling Filter
T.S.	Total Solids
V.S.	Volatile Solids
>	Greater than
<	Less than
Wt.	Weight

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